HEAT INTEGRATION AND RETROFITTING ANALYSIS OF ARABIAN LIGHT CRUDE DISTILLATION UNITS

BY

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DEDICATION

I would like to dedicate this academic project to my beloved parents, who encouraged me and supported me to continue my education and helped me to purse my educational objectives and set my essential goals in my life.

Special thanks are due to my wife who has always been there for me supporting and encouraging me with unconditional love to continue my education and to make my dreams come true. Also, much love to my daughters Raqia and Reem, the lights of my life. I would like to thank all my brothers and sisters and the rest of my family for their love and support.

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LIST OF ABBREVIATIONS

A	:	Heat Transfer Area (ft^2)
CP	:	Specific Heat Capacity (MMBtu/hr ^{0}F)
ΔH	:	Change in Flow Enthalpy (MMBtu/hr)
K	:	Number of Temperature Intervals or Segments
m	:	Mass Flowrate (Ib/s)
QH,min	:	Minimum hot utility requirement (MMBtu/hr)
QC,min	:	Minimum cold utility requirement (MMBtu/hr)
HE	:	Heat Exchanger
HEN	:	Heat Exchanger Network
HRAT	:	Heat Recovery Approach Temperature (^{0}F)
LP	:	Linear programming
T	:	Temperature (^{0}F)
TS	:	Supply Temperature of Process Stream (^{0}F)
TT	:	Target Temperature of Process Stream (^{0}F)
ΔT	:	Temperature Difference (^{0}F)
ΔTmin	:	Minimum Allowed Temperature Difference (^{0}F)
HS-1,2..	:	Hot Stream Number
CS-1,2..	:	Cold Stream Number
XP	:	Across-Pinch

CDU	:	Crude Distillation Unit
VDU	:	Vacuum Distillation Unit
LPG	:	Liquid Petroleum Gas
LDO	:	Light Diesel Oil
HDO	:	Heavy Diesel Oil
HDGO	:	Heavy Diesel Gas Oil
LVGO	:	Light Vacuum Gas Oil
HVGO	:	Heavy Vacuum Gas Oil
R1 to R58	:	Variables Represent Heat Residual (MMBtu/hr)
hu	:	Hot Utility Exchanger
cu	:	Cold Utility Exchanger
A,B,C	:	Heat Exchanger Cost Parameters

ABSTRACT

Full Name : Badiea Saeed Omer Babaqi

Title of Study : Heat Integration and Retrofitting Analysis of Arabian Light Crude Distillation Units

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The heat integration and retrofit analysis of Arabian Crude Distillation Units was carried out to identify opportunity for energy savings using different design options, to arrive at optimum heat exchanger network. Data used were extracted from existing and operating crude oil refining plant in the Kingdom of Saudi Arabia (KSA). The Pinch Analysis of existing plant using process integration software (Heat-Int) revealed that the hot and cold utilities consumptions at the prevailing ΔT_{min} of 77^{o}F in the plant were 680.23 MMBtu/hr and 521.215 MMBtu/hr, respectively. Economic evaluation of existing plant revealed total operating cost of $4,829,625/yr. Retrofit of existing plant using different design options generated an optimal network comprised four additional heat exchangers and repiping of one existing exchanger with reduction in ΔT_{min} from 77^{o}F to 57^{o}C. The hot and cold utilities consumptions also reduced to 623.455 MMBtu/hr and 464.437 MMBtu/hr, respectively. Economic evaluation of the new design showed energy savings of $259,860/yr. The trade-off between energy and capital cost in the procumbent of new exchangers and installation gives overall cost savings of $3,280/yr with payback period of one year. The study concluded that retrofitting of existing heat exchanger network to

correct those exchangers that are transferring heat across the pinch could result in enormous energy savings.

ملخص الرسالة

الإسم : **بديع سعيد عمر بابقي**

عنوان الرسالة : **التكامل الحراري والتحليل التحسيني لوحدات التقطير للنفط العربي الخفيف**

التخصص : **هندسة كيميائية**

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تم تنفيذ التكامل الحراري و التحليل التحسيني للوحدات التقطير للنفط العربي الخفيف من أجل تحديد الفرص لتحقيق وفورات الطاقة باستخدام خيارات التصميم المختلفة ، وللوصول إلى أفضل شبكة مبادلات الحرارية . تم استخراج البيانات المستخدمة من مصنع تكرير النفط الخام الموجود والعامل في المملكة العربية السعودية (KSA) . تحليل القرصة للمصنع الموجود باستخدام برامج عملية التكامل (Heat-Int) الذي كشفت أن كمية الاستهلاك للمرافق الساخنة والباردة عند 77 درجة فهرنهايت في المصنع كانت 680.23 مليون وحدة حرارية / ساعة و 521.215 مليون وحدة حرارية / ساعة على التوالي . كشف التقييم الاقتصادي للمصنع القائم إجمالي تكاليف التشغيل حوالى 4829625 دولار / سنة. التحديث للمصنع الموجودة باستخدام خيارات التصميم المختلفة حيث تم إنشاء شبكة مثلى تتألف من أربعة مبادلات حرارية إضافية و تحويل مبادل موجود الى موضع آخر مع إنخفاض في درجة الحرارة من 77 درجة فهرنهايت إلى 57 درجة فهرنهايت . كما خفضت استهلاك المرافق الساخنة والباردة إلى 623.455 مليون وحدة حرارية/ ساعة و 464.437 مليون وحدة حرارية / ساعة على التوالي . وأظهر التقييم الاقتصادي للتصميم الجديد توفير الطاقة من 259860 دولار / سنة. المفاضلة بين الطاقة وتكلفة رأس المال في المبادلات جديدة وتركيب يعطي فورات في التكاليف الإجمالية 3280 دولار / سنة مع فترة الاسترداد سنة واحدة. و بينت الدراسة أن التعديل التحديثي القائمة لشبكة المبادلات الحرارية أن تصحيح تلك المبادلات التي يتم نقل الحرارة عبر قرصة يمكن أن يؤدي إلى وفورات في الطاقة الهائلة.

CHAPTER 1

INTRODUCTION

1.1 General Introduction

Petroleum refining is a process of separating crude oil to various fractions of primary and secondary fuels and petroleum products, which help better the quality of life. The world relies largely on energy, where the importance of energy has become essential to reduce its consumption. Crude oil is the first alternative source of energy, so energy saving investment as well as environmental protection from emissions were the main aim of investment that make more attractive as fuels become more upmarket.

The increasing cost of energy and environmental impact of its combustion has necessitated the operation of energy intensive processes in the most efficient and economical manners. Refining operations are major part of an economy, and efforts are continually been made to improve their performance in order to provide solutions to energy utilization problems and reducing emissions of greenhouse gases and byproducts of combustion from the industry. Since fuel constitutes a large percentage of the operating cost in any process industry, all efforts towards minimizing its consumption correspond not only to reduction of hazardous gases emissions but also increase profitability associated with energy savings. [2, 16, 42]

Energy conservation opportunities within petroleum refineries have remained one of the most significant areas of research today [7]. This is because refinery system comprises series of process streams that require heating and cooling through inter connection of heat exchanger network and external utilities such as steam, cooling water, cooling fan etc. Therefore, the performance of heat exchanger network in the conservation of energy resources in such an energy intensive process cannot be over emphasized. One of the successful techniques for investigating energy integration and efficient design of heat exchanger networks is pinch technology. Pinch Analysis uses thermodynamic concepts and heuristics. The concepts of pinch analysis have been applied to design of new plants with reduced energy and capital costs and for improving existing processes to ascertain efficiency and provide potential design modifications to improve performance. [16, 17]

Management technology of energy can be achieved either by paying less per unit of energy by co-generation of heat and power, or reducing energy consumption per unit of product via energy integration of process plants. Because of energy technology, the industrial sectors have met enormous challenges about environment protection from emissions and the output of high quality product since discovery of crude oil in 19th century. Refining of crude oil worth's about 50% of cash operating costs on energy (without capital costs and depreciation), producing energy a main cost factor and also an essential chance for cost reduction. Energy use is also a major source of emissions in the oil refining industry making energy efficiency enhancement a smart opportunity to decrease emissions and operating costs. [6, 19, 39]

Management of the energy conservation and its efficiency has become a vital objective of all energy intensive industries. Nowadays, a lot of energy consumed by the industrial sectors, especially petroleum refinery industries. For solving this problem, the energy consumption has to be cut down; therefore, exploring energy conservation opportunities within petroleum refineries have remain the most significant area of research in many refining and petrochemical industries. Crude and vacuum distillation units (CDU & VDU) are the main source of energy saving in oil refineries due to their characteristic of high energy consumption along with larger influence on the overall operability and profitability of the refinery complexes. Studies have shown that optimization of utilities, heat exchangers networks (HEN) and fired heaters offers substantial energy improvement and low-investment opportunities. [11, 39, 40]

The crude oil distillation (CDU&VDU) is the most important process in the oil refinery because it produces a wide range of products such as gasoline, naphtha, kerosene, diesel, etc., at same time it is greatly energy consumption, where energy is consumed often as direct and indirect fuel in petroleum refineries. The direct fuel used in the process heaters, while the indirect fuel used for raising steam for generating power.

Conservation in oil industry may mean making new investments via applying of technologies or in more efficient equipment. There is a need for new processes that either increase the yields of higher value products or simply are much energy efficient or use low value fuels and products such as those that use less severe operating conditions. Furthermore, integrated approach towards system design and operation to exploit full synergy of the processing site are needed. Then, the choice of type of energy (e.g. refrigeration or cooling water, LP steam or MP steam) and waste heat recovery can be substantial source of energy savings.

Oil refineries have large of energy-intensive process especially crude distillation unit (CDU) and one the vital area for heat integration. Heat exchangers network of the distillation unit consists of a crude pre-heat exchanger network and flashing section atmospheric and vacuum distillation section. Moreover, the modification process of heat exchanger network often added new heat exchanger or more exchanger area to existing heat exchanger network, or relocating the heat exchangers. [19, 22, 28]

General heat integration of refinery utilities and heat exchanger network is very important for energy saving and cost reduction. Energy conservation opportunities within petroleum refineries have remain a most significant area of research in many refinery and petroleum industries. Due to more locations have waste energy in refinery sector such as loss of stack gases, lack of process integration, water-cooling stream etc. One of the methods may improve conservation of energy revamping the existing unit for design modification. Furthermore, units of waste heat recovery are pre-heaters, convection section and steam generators in furnaces.

Refinery often used fuel in process heaters or boiler about 80% of energy consumption. The main area of heat waste in combustion process is the stack gasses exiting the furnaces are high temperature. [7, 26]

In the last thirty years, syntheses of the heat exchanger network and heat integration have been an important investigation activity in the process systems. The total costs minimization in heat exchanger network and retrofit designs determined by using mathematical program in 1995. [8]

Rising energy costs have forced many refineries to consider the interest in heat exchanger networks retrofit. However, the first organized approaches to retrofitting we are not presented until the mid-1980s. Early methods were based principally on synthesis methods for grass-root design. Jones et al. used a commercial program to generate grass-root designs, and then selected the design closest to the existing one for further development. Saboo and Morari used a similar concept but a more rigorous technique than Jones et al. The major difference between grass-root design and retrofit design is the amount of constraints forced on the solution by the existing layout of the process in the retrofit location. [30, 15, 27]

The major products of the plant include liquid petroleum gas (LPG) production, stabilized whole naphtha, kerosene, heavy kerosene, light diesel oil (LDO), heavy diesel oil (HDO), heavy diesel gas oil (HDGO), light vacuum gas oil (LVGO), heavy vacuum gas oil (HVGO) and vacuum residuum.

In this study, pinch technology and retrofitting analysis is applied on the existing heat exchangers network of the plant for saving of energy consumption and reducing operating costs for obtaining the best design which have maximum heat recovery. Economic constraints such as investment and payback are considered.

1.2 Significance of Study

The importance of heat integration of the crude distillation plant is essential process due to energy intensive. Therefore, the high energy consumption with larger influence on the overall operability and profitability will become not economical of the plant. So energy saving as well as environmental protection from emissions was the main aim of investment.

This study applied pinch analysis technology for saving of energy consumption and reducing operating costs, likewise, for environmental protection from emissions.

1.3 The Objective of Study

The objectives of this research can be summarized as follows:

1. Determining the minimum energy targets for the heat exchanger networks of crude and vacuum distillation units in oil refining industry.
2. Optimization of utilities types and quantities used in the plant.
3. Solving for optimum heat exchangers network.
4. Retrofitting of the existing heat exchangers network towards the optimal design.

CHAPTER 2

LITERATURE REVIEW

2.1 Introduction

In any process chemical consider the design of heat exchangers network is very important. The chemical process design in any plant illustrates a hierarchical nature. In figure 2.1, the layers represent overall design of any process where the design of process starts at the reactor that converted the feedstock into product. The second layer is the separation system incudes once the feed, product, recycle, concentration and flow rates. The reaction separation layer establishes the process mass and heat balance, which dictates the heat exchanger network requirement. The last layer represent heating and cooling energy requirement that is not recovered by process streams is satisfied by use of utilities.

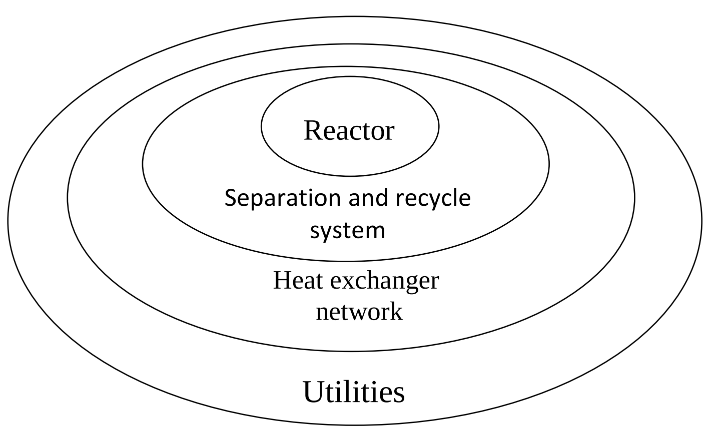


Figure 2.1: Process Designing Onion Model

2.2 Review of Oil Refining

The petroleum refining processes are extremely complicated and integrated. Heat integration of refinery utilities and heat exchanger networks (HENs) using pinch technology was first discussed by Linnhoff et al 1983, since then, numerous investigation of refinery crude and vacuum units have been carried out using pinch methodology [(3,29,33)]. One of the main advantages of this method is that it provide pathways for identifying trade-off between energy and capital cost besides minimize utility and energy consumptions [(30,31)]. Fraser and Gillespie (1992) used pinch technology method of all the units of an existing oil refinery for improving heat recovery efficiency. They obtained 22% saving for energy recovery target while current usage 16% saving. The way has done changing the matching arrangements of network and adding extra area on reactor feed and effluent heat exchangers. [(9)] Integrated approach was developed in 1999 by Suphanit to reduce energy of crude oil distillation system design. He applied shortcut distillation models, optimization framework and pinch analysis and, to make more energy efficiency grassroots design. [(31)]

Al-Riyami et al. (2001) studied heat integration for fluid catalytic cracking plant to improve energy recovery and performance of the existing heat exchanger networks. They found targeting stage by the incremental area efficiency method and retrofit design of heat exchanger network by the network pinch method. The current network had a ΔT_{min} 24^oC and the area efficiency about 80.5%. The minimum ΔT_{min} of the existing design (11.5 ^oC) for target stage using the incremental area. The results were 27% utility cost savings. The optimal modification consists of the repiping of one existing exchanger and adding four new heat exchanger units. [(3)] Furthermore, Al-Mutairi (2010) studied optimal design of heat exchangers network for energy savings to reduce of operation cost. He used Aspen simulation software and mathematical programming method (MILP) in order to optimize the production and minimize the hot and cold utilities. [2]

Rodera et al (2001) used pinch analysis method in the crude distillation unit. They obtained maximum savings of energy and economic analysis identified a retrofit chance with a payout of 1.2 years. [25] On the other hand, Querzoliet al (2003) studied crude distillation and residue cracking units for improvement the energy efficiency of the existing networks. They used targeting before design approach, where they found reducing crude distillation unit requirement by 40% for new grass roots design at ΔT_{min} 55^{o}C. [24]

Yu-lin (2011) studied crude distillation unit and carried out analysis of heat exchangers where he investigated energy saving in heat exchangers network using pinch analysis method. He discovered through his study of the grid of HEN that number of heat exchangers without splitting of stream is more than within splitting of stream. [41]

The importance energy saving in oil refining has become necessary for reducing the cost; where Ballut (1986) analyzed crude preheat train in distillation unit to raise heat recovery of process via adding new heat exchanger to increase crude inlet temperature to the heater. [5] Papalexandri' and Patsiatzis (1998) studied a crude preheat system of crude distillation unit to energy reduction of consumption by using heat integration aspects. They carried out balance between production of steam and consumption of fuel. [21] Moreover, Chegini et al. (2008) investigated modification of preheating in crude distillation unit. They used Heat-int Software for heat integration of the network with the aim of heat load reduction in the furnace. The results were 9855kW heat duty of distillation furnace that loads to saving about 820,000 US$/year and decreasing the temperature of furnace flue gas to 290 ^{o}C and green hose emission to 20,000 tones/year. [6] In 2009, Ajao and Akande used pinch analysis in crude distillation unit, where they carried out heat integration of crude preheat train for maximum energy recovery via additional heat exchangers. [1]

Several design methods for the heat exchanger network retrofit have been suggested during the three decades. Gadalla et al. (2003) studied refinery distillation systems to decrease energy-consumption where used retrofit shortcut model for the heat exchangers network; including existing exchanger matches, existing areas and duties. In addition, shortcut model for the simulation of the existing distillation column. They showed that a economizing in energy consumption and operating costs about 25%. [10]

In 2007, Shanazar et al. also studied retrofit of crude distillation unit for energy keeping by using integrated tool. They used Aspen Plus and Aspen Pinch to get grassroots design of heat exchangers network. They found the best heat recovery at furnace about 9.246% of overall energy consumption. The modifications have some changes; such as added new heat exchanger, split of stream and repiping existing heat exchanger. [28]

In addition, Promvitak et al. (2009) studied retrofit of heat exchanger networks of CDU. They can cut down energy at furnace and coolers after additional exchanger area for preheating crude when applied pinch method and stage model. They found that energy-saving about 1.3% at furnace and 2.8% at coolers after retrofitting of heat exchangers network for CDU. [22]

Wang et al. (2011) studied heat exchanger network retrofit by heat transfer improvement via addition shell side or tube-side or both, which depends on the heat transfer coefficient. This method can save energy in retrofit design without any topology modification due to avoid increasing capital cost for replacing and pipelines. The method of optimization based on simulated annealing to obtain the suitable heat exchangers. [38]

2.3 Concepts of Pinch Technology

Pinch technology is rigorous method of heat exchanger network design and it is based on thermodynamic principles. The composite curve and grand composite curve are a fundamental established concept in the pinch analysis. The composite curve is relationship between temperature and enthalpy of the process streams. Figure 2.2 shows composite curve that include hot composite and cold composite separated by a minimum temperature difference (ΔT_{min}). For feasible heat exchange, the hot composite is always above cold composite which according to laws of thermodynamic i.e. heat is transferred from a hot body to a cold body. The process streams data that consist of supply and target temperatures and heat capacity flow rate, construct the composite curves.

The pinch separates the process into heat sink that locates above pinch and heat source that locates below pinch, where the process is in heat balance with both minimum hot and cold utilities, respectively. Therefore, to achieve minimum utilities requirement in the process design, there are three golden rules: no external heating below the pinch, no external cooling above the pinch and no heat transfer across the pinch.

The grand composite curve is graphical tool that is used in the pinch analysis and it is a plot shifted temperature versus enthalpy such as figure 2.2. It shows the net heat deficient or surplus at temperature intervals and it gives the duty and temperature information that is required by the process.

Pinch technology is a vital tool that is applied engineers a practical way to investigate and design chemical processes and made it user-friendly to integrate plant design by reordering equipment, for instance, reactors, separators, distillation columns, pumps and heat exchanger networks.

It is used to improve the design for complicated processes with unit operations and several utilities. The method included fixing the level and enthalpy change via use composite curve and beside grand composite curve at setting the low cost for many different utility levels. [35]

Hallale (2001) studied process integration, which is the main part of the application of this pinch technology. Process integration uses to various applications such as process operations, energy efficiency, efficient use of raw materials and emission reduction. He used and investigated in column analysis and integration and design of heat exchanger networks and its retrofit etc., where he began from creating composite curve then finding pinch point, putting targets and final structure of the modifying of heat exchanger network to make the minimum target of utilities.[13]

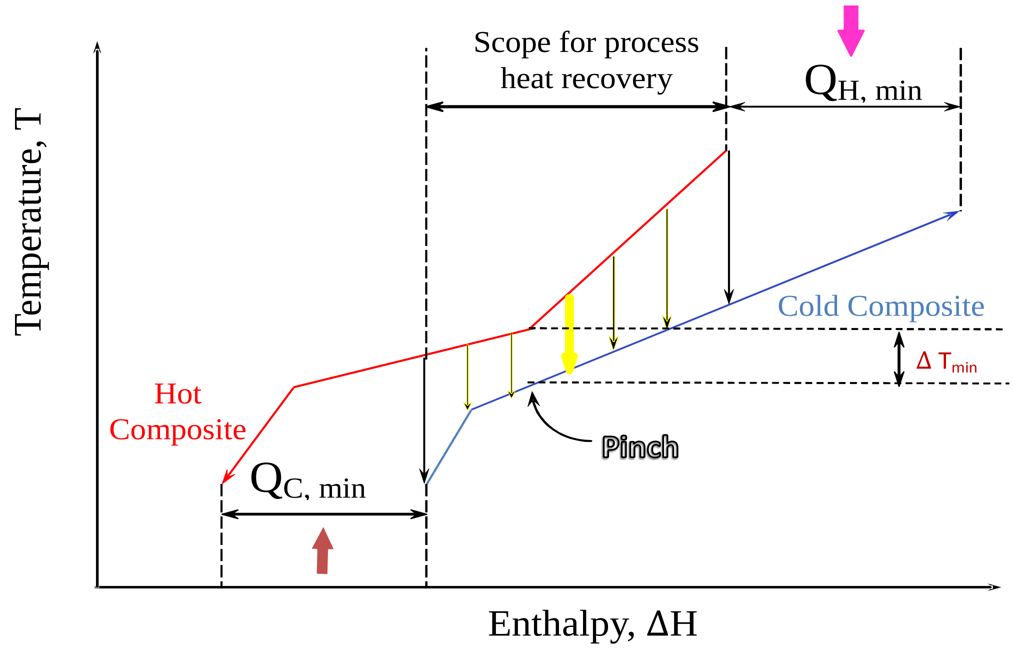


Figure 2.2: Composite Curves for Pinch Analysis

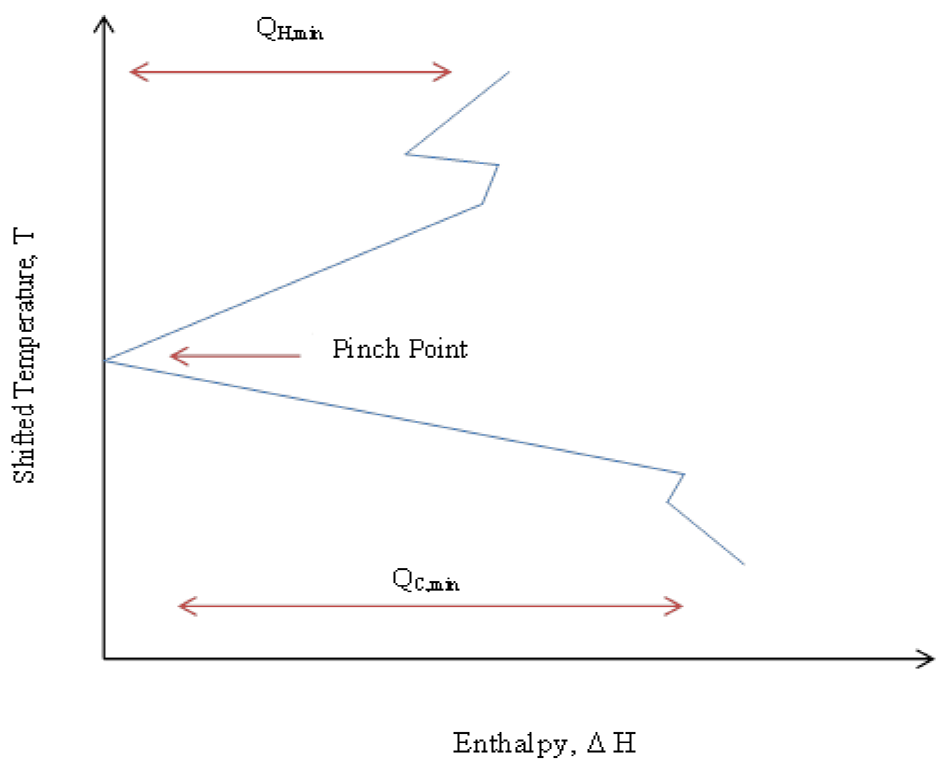


Figure 2.3: Grand Composite Curve for Pinch Analysis

2.4 The Pinch Design Method for HEN Design and Retrofit

The pinch design method was first design methodology by Linnhoff and Hindmarsh in1983. The design of the network has to achieve of the target, besides satisfying the pinch principles and the feasibility conditions at the pinch by stream splitting. The tick-off heuristic is a good way but this can correct the energy usage. By using heat load loop and heat load path for enhancing the design of heat exchangers network and taking account of balance between costs of energy and capital. (18)

In 1990, Linnhoff and Ahmad studied design of heat exchanger networks with trade-off between cost of energy and capital. By capital cost model, the methodology was created location cost targets and optimizing to the design.(17)

The method used in pinch design for retrofitting of process was presented, by Tjoe and Linnhoff (1986). This method assumed that retrofit in heat exchanger network will find the best network of the process. Initially, we use the area-energy curve to set the target. The energy saving and minimum temperature are set under an identified investment or payback time. The retrofit is to recognize or revamp the cross of the pinch exchangers (XP) and modify them. (34)

The method for retrofitting of heat exchanger networks were presented by Polley and Amidpour (2000). They selected the payback time and capital investments as the main economic indicators for process retrofit. To establish the retrofit target, they used the saving-investment plot. The retrofit analysis existed by comparison of the area efficiency between the performance of the unit with the ideal. The cross-pinch exchangers were recognized and then modified using in a conventional method. They suggested the new technique by recognizing the structure of the units in the first step after that the trade-off of energy-investment will be made to size and modify the exchangers.[23]

2.5 Process Heat Integration

Heat integration means incorporating different processes to achieve of energy conservations. Heat integration methodology is established on the understanding and analysis of streams of the process via exchanging between them by using a temperature-enthalpy diagram.

There is the integration of heat engines, distillation columns, heat pumps, with the processes to achieve more advantages. Townsend and Linnhoff (1983a) investigated of placement of heat pumps and engines in networks of the process. The appropriate placement is the position at which we can get benefits over the stand alone heat engines or heat pumps. For the heat engines, they indicated that an appropriate placement is not across process pinch while the suitable placement is above or below the pinch. On the other hand, appropriate placement of heat pumps is contrasting. The suitable placement is in across the pinch. In addition, they said that cannot achieve a completely applicable integration because of the heat has to cross-pinch to the ambient.[36]

In 1983, Townsend and Linnhoff put systems of the design for alternative the best practical technology via using of the conditions planned. The process of heat source and heat sink was introduced in this procedure.

Practical design constraints are always taken into account in this method (pinch analysis). Also, it included power generation and heat recovery via integration for chemical processes and others, where it used to assess options and to classify preferred structure at the initial step of design. [37]

CHAPTER 3

PROCESS DATA

3.1 Introduction

The main things in a process integration study of the plant are mass and heat balance of the process, which gives information about the current operation and the heat exchangers network, thermal data and economic data of the process. The data extraction incudes all hot and cold streams of the flow diagram of the crude plant.

The refinery under study at present is the oldest and biggest of the refineries in Saudi Arabia. Although the refinery came into live in 1945 with a capacity of 50,000 barrels/day, it has however subjected to several revamping and upgrading processes in an attempt to increase energy efficiency and processing capabilities of its various units.

The data needed for the heat integration of the distillation units heat exchanger is extraction data, cost and economic data.

3.2 Description of the Plant

The plant includes four sections that are: atmospheric section, vacuum section, naphtha stabilizer section and asphalt section. The products from this plant are: LPG, stabilized whole naphtha, kerosene, Diesel gas oil (DGO 2, 3, 4) combined, heavy diesel gas oil

(HDGO), light vacuum gas oil (LVGO), heavy vacuum gas oil (HVGO) and vacuum residuum.

The atmospheric column receives the crude charge and separates it into overhead product, kerosene, diesel DGO, and reduced crude. The naphtha stabilizer receives the atmospheric overhead steam and separates it into LPG and stabilized naphtha. The reduced crude is charged to the Vacuum Tower were it is further separated into heavy diesel (HDGO), light vacuum gas oil (LVGO), heavy vacuum gas oil (HVGO), and vacuum residuum. The vacuum residuum can be charged to the asphalt section.

Figures (3.1 to 3.10) show a simplified overall view of the plant. When the crude received from the tank farm, the crude pumps by G-101A/C to the heat exchanger E-101A/D where the crude exchanges the heat with the overhead stream from the atmospheric column (C-100) then the crude goes to a primary preheat exchangers. The primary heat exchangers consist of five heat exchangers. In the primary heat exchangers the crude split into two streams in parallel. The upper stream has two heat exchangers E-261A/C where the crude exchanges the heat with VGO which is side cut six products from the vacuum column and E-131 where the crude exchanges the heat with the diesel which is side cut three products from the atmospheric column. The downstream has three heat exchangers which are E-142 where the crude exchanges the heat with the diesel which is side cut four products from the atmospheric column, E-121 where the crude exchanges the heat with the diesel which is side cut two products from the atmospheric column and E-116A/C where the crude exchanges the heat with the atmospheric top circulating reflux.

After that the two crude streams recombine together and the crude goes to the desalter (D-101). The crude unit passes through a primary preheat exchanger before entering the desalter in order to increase the crude temperature to the optimum temperature for desalting. The desalter removes contaminants (salt, solids and entrained water) from the crude. After that, the crude enters a flash drum in order to separate the gas from the liquid where the overhead gas enters the atmospheric distillation column (C-100) in stage 5 and the liquid pumps by G-102 A/C to a secondary preheat exchangers.

The secondary preheat exchangers consist of nine heat exchangers and the crude split into two parallel streams. The upper streams has four heat exchangers which are E-283A/B where the crude exchange the heat with HVGO which is side cut eight product, E-271A/B where the crude exchanges the heat with LVGO which is side cut seven product, E-292A/F where the crude exchanges the heat with vacuum residuum which is the vacuum bottom product and E-281A/C where the crude exchanges the heat with HVGO which is side cut eight product. The downstream has five consecutive heat exchangers which are E-115A/C where the crude exchanges the heat with the atmospheric top circulating reflux, E-141A/B where the crude exchanges the heat with the diesel which is side cut four products from the atmospheric column, E-282A/C where the crude exchanges the heat with HVGO which is side cut eight product, E-145A/C where the crude exchanges the heat with the atmospheric bottom circulating reflux and E-291A/C where the crude exchanges the heat with vacuum residuum which is the vacuum bottom product. After that the two streams recombine in the end of the secondary preheat exchangers and then its splits again to enter furnaces F-100A/B to rises the temperature to the optimum for the distillation. Next the liquid crude recombines again and enters the atmospheric column in stage 6.

When the crude enters to the atmospheric column (C-100) the packing beds reduce the differential pressure needed for product separation and improve distillation. Atmospheric column products are: Whole naphtha (the lightest component, the overhead product), kerosene (side-cut 1product), diesel (side-cuts 2, 3 and 4 product) and reduced crude (bottom product).

The whole naphtha is a part of the atmospheric overhead stream which goes to E-101A/D to become cooler when it exchanges its heat with the crude. After that the overhead stream goes to D-124 which uses for preventing the corrosion issue in atmospheric column head. Then the stream passes through the fin fan E-102A/N and then it goes to D-103 where the stream separates into three parts. The first part is the off gas. The second part is hydrocarbon which goes to two different paths the first path pumps the hydrocarbon by G-301A/B to naphtha stabilizer water coalesce (D-301) where the second path pumps the hydrocarbon by G-103A/B as C-100 reflux. The third part is the water which pumps by G-106A/B to two paths the first path is to the seal drum (D-201) and the other path as C-100 quench.

In D-301 the hydrocarbon separates from water and then goes to E-301A/D to exchange the heat with the naphtha stabilizer column product to rise its temperature to the optimum one to help separating the naphtha from the LPG (C4+ light ends) in the naphtha column (C-300). The hydrocarbon feed enters above tray 20. The hydrocarbon flows to the C-300 in order to separates the naphtha from the LPG. The LPG flows from the top of the stabilizer as overheads and its goes to the reflux drum (D-300) either directly or by pass through fin fan E-300A/B. In D-300 the LPG separated from water and pumps by G-300A/B and its split either as a reflux stream or as a product. The water from D-300 is send to D-201 and the off gas is send to plant 410. The stabilized naphtha flows from the bottom of the stabilizer. Then the stabilized naphtha flows to a heat exchanger E-301A/D to cool it. After that the stabilized naphtha flows to the fin fan E-302A/C and then it goes to plant J24.

The kerosene which is side cut one product its flows in two paths. The first path is the atmospheric top reflux which goes to G-115A/B and pumps the kerosene to E-115A/C and E-116A/C where the kerosene exchange its heat with the crude and then it flows to the fin fan E-117A/B after that it return to the column. The second path it flows to the stripper C-110 which strip the light material by the 60 psig steam and send it back to the column from the top whereas the stripper bottom stream pumps by G-110 to either to the DHT or to fin fan E-112A/B then to the tank farm.

The diesel from side cut two flows to the stripper C-120 which strip the light material by the 60 psig steam and send it back to the column from the top whereas the stripper bottom stream pumps by G-120A/B to E-121where the diesel exchange the its heat with the crude and then its flows either to the DHT or to fin fan E-122A/B and then to the tank farm.

The diesel from side cut three flows to the stripper C-130 which strip the light material by the 60 psig steam and send it back to the column from the top whereas the stripper bottom stream pumps by G-130A/B to E-131where the diesel exchange the its heat with the crude and then its flows either to the DHT or to fin fan E-133 A/B and then to the tank farm.

The diesel from side cut four product flows in two paths. The first path is the atmospheric bottom reflux which goes to G-145A/B and pumps the diesel to E-145A/C where the diesel exchanges its heat with the crude and then it flows to E-305 where the diesel exchanges its heat with the naphtha and then it splits into two streams the first goes directly to the column and the second passes through E-146A/C where it exchanges the heat with circulate boiler water and then it enters the column. The second path flows to the fin fan E-117A/B after that it returns to the column. The second path it flows to the stripper C-140 which strips the light material by the 60 psig steam and sends it back to the column from the top whereas the stripper bottom stream pumps by G-140 A/B and it flows into E-141 A/B and E-142 where the diesel exchanges its heat with crude and then its flows either to the DHT or to fin fan E-143A/B and then to the tank farm. The side cut five stream is used as a pump around. The atmospheric bottom product is the reduced crude and it pumps by G-160A/B to vacuum section.

The vacuum section receives its feed from the bottom product of the atmospheric column (C-100). When the reduced crude enters the vacuum section it splits into two streams and the two streams enters F-200A/B in order to increase the temperature of the reduced crude to the optimum temperature for the vacuum distillation. The partially vaporized reduced crude leaves the furnaces at 745-770 ^{o}F and flows to the vacuum column (C-200).The vacuum distillation products are: Vacuum residuum (the heaviest hydrocarbon, bottom product), vacuum gas oil (VGO, side cut 6 product), light vacuum gas oil (LVGO, side cut 7 product), heavy vacuum gas oil (HVGO, side cut 8 product).

The vacuum gas oil (VGO) pumps by G-260A/B to E-261A/C where the VGO exchanges its heat with the crude then the VGO splits into three paths. The first path is goes to fin fan E-262A/D and then it flows back to the vacuum column as a vacuum top circulating reflux. The second path flows directly to the DHT. The third path flows to fin fan E-263A/B and then it flows to the tank.

The light vacuum gas oil (LVGO) pumps by G-270A/B to E-271A/B to exchanges its heat with the crude and then it splits into three paths. The first path flows to J-80 directly. The second path to E-293 where the LVGO mixed with the vacuum residuum and exchanges its heat with the boiler feed water and then it flows to J-85. The third path flows to fin fan E-272A/B and then it splits into two streams, one stream flows to the tank and the other flows to J-80.

The heavy vacuum gas oil (HVGO) splits into two streams. The first stream pumps by G-280A/B to E-281A/C and then to E-282 A/C to exchange its heat with the crude, then its split to two paths, the first path join the mid reflux stream after G-275A/B and return to the column, the second path flows to E-283A/B where it exchanges its heat with the crude and then it goes to the tank. The second streams it returns back to the column as wash oil.

The vacuum overhead stream flows to E-201A/D where it exchanges its heat with the sea water. The gas moves up to the ejector (K-201A/D) where the gas sends to E-202A/B and the condense liquid flows to the seal drum (D-201). E-202A/B receives the gas from K-201A/D where the gas exchanges its heat with the sea water. The condense liquid flows to D-201 and the gas goes to K-202A/B which send the gas to E-203A/B where the gas exchanges its heat with the sea water. The condense liquid flows to D-201 and the gas goes to K-203A/B which send the gas to E-204A/B where the gas exchanges its heat with the sea water. The condense liquid flows to D-201 and the gas goes to K-204A/B which send the gas to the atmosphere. The seal drum D-201 also receives the condensate from D-103, D-300 and D-301. D-201 separates oil from the water where the oil sends by G-201A/B to the crude feed suction and the water sends by G-202A/B to either G-101A/C discharge, to plant 11 or to desalted water break tank in plant J-37. The waste gas sends to furnaces or to relief.

The vacuum residuum which is the vacuum bottom product pumps by G-290A/C to E-291A/C and then to E-292A/F where the vacuum residuum exchanges its heat with the crude. After that the vacuum residuum sends to the asphalt facilities.

When the asphalt section receives the vacuum residuum from the vacuum section, the feed enters to E-501A/C and exchanges its heat with the boiler feed water. Then it enters to the paving asphalt converter C-501 where it mixes with quench steam, purge steam and air. The asphalt converter product flows from the bottom to G-501A/C which pumps the asphalt either to the other paving asphalt converter (C-503) or to the mixer KM-502A/B which mix the asphalt with the vacuum residuum and sends it to E-502A/F where the asphalt exchanges its heat with the boiler feed water and then the asphalt flows to the tank. The waste gas flows from the top of the converter to the knocks out drum (D-501).

The second paving asphalt converter (C-503) receives its feed from C-501 bottom and mix it with quench steam, purge steam and air. The waste gas from C-503 flows to the waste gas drum. C-503 product flows from the bottom to G-503A/B which pumps the asphalt either to the reflux stream or to E-503A/D where the asphalt exchanges its heat with boiler feed water and then it flows either to the reflux stream or to the tank.

The steam drum (D-500) receives the boiler feed water and circulates it through E-501A/C, E-502A/F by G-500A/C and E-503A/D. The steam knocks out drum (D-501) receives the waste gas from C-501 and C-503 and sends it to the furnace F-500. The steam knocks out drum (D-502) receives the 60 psig steam. The overhead product sends as 60 Psig quench steam to C-501 and C-503, 60 psig purge steam to C-501 and C-503 and to the furnace F-500 as a waste gas. The bottom product flows as LP condensate.

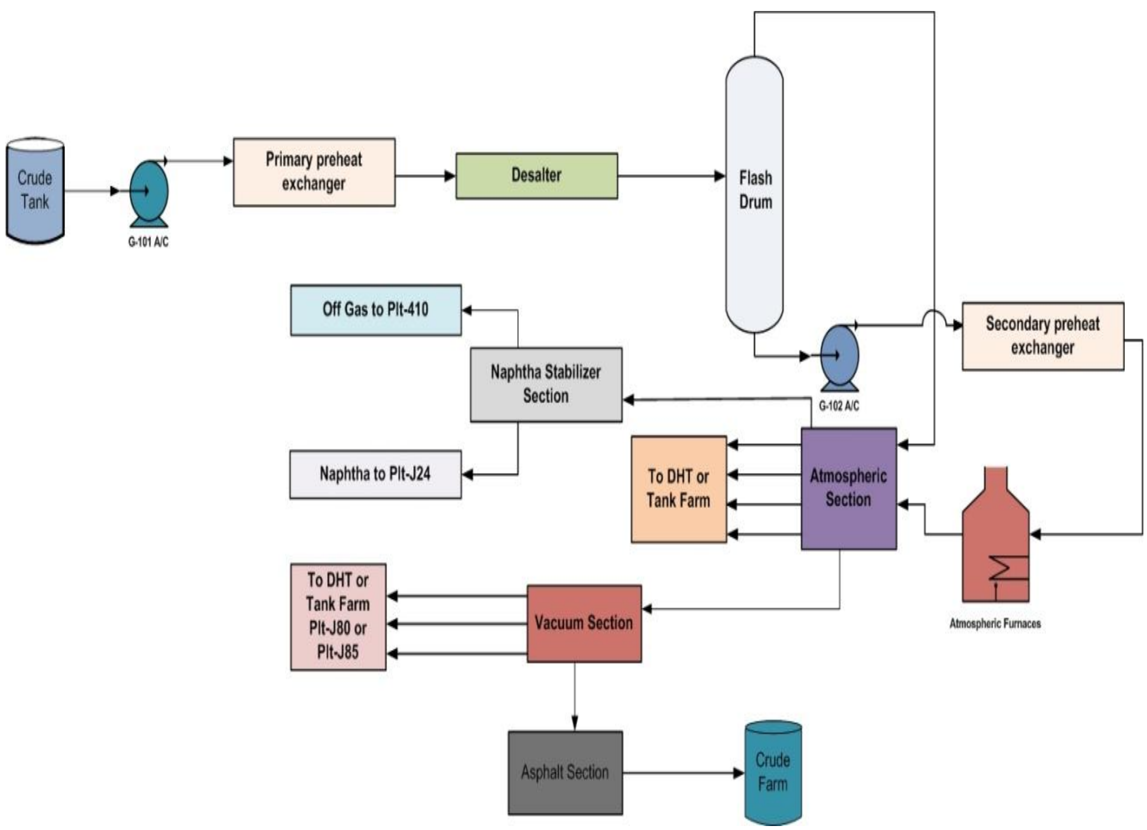


Figure 3.1: Schematic of a Typical CDU& VDU Process

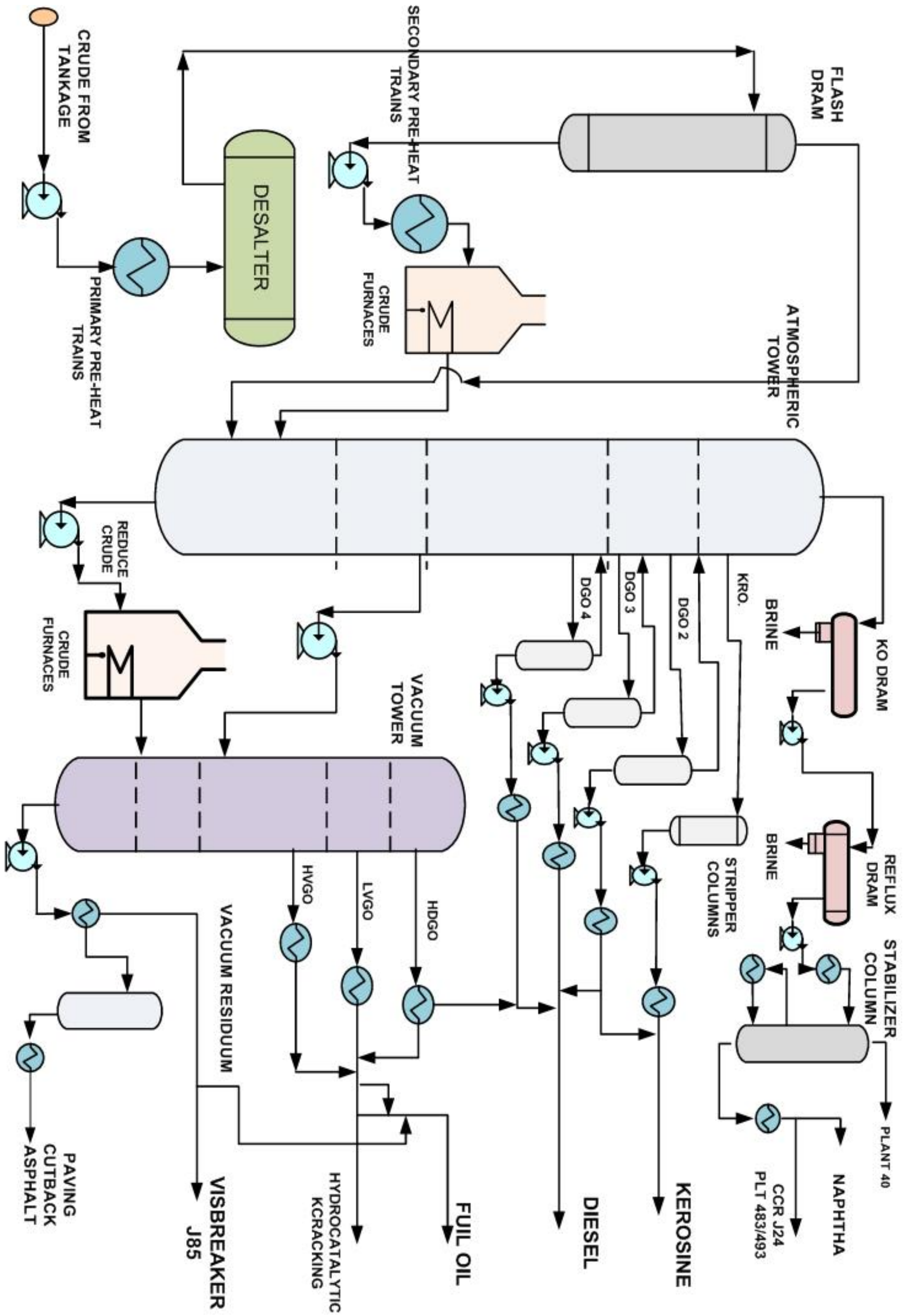


Figure 3.2: The Crude Distillation Unit

3.2.1 Crude Preheat Description

Crude feedstocks such as Arabian Light (AL), Arabian Extra Light (AXL), and Arabian Medium (AM) flows from a storage tank farm to the atmospheric overhead/feed condenser E-101 A-D (Figure 3.3)

They are heated prior to their separation into two streams (trains A and B) which flows to primary crude preheat exchangers. Inside the exchangers, the crude is heated to their desalinating temperature of about 280 – 310 ^{o}F. As the two crude streams leave the primary crude preheat exchangers, they recombined and flow into the desalter D-101 which desalinate the crudes as shown in Figure 3.3.

The desalted crude flows from the top of D-101 to flash drum D-102 were they are separated from the light ends (vapor). The vapor upon separation flows to the atmospheric column flash zone through the top most section of the desalter. On the other hand, the preflashed crude flows from the bottom of D-102 to the secondary heat exchangers train were they are further separated into two streams before moving to the secondary crude preheaters. Upon leaving the preheaters the crude recombined before separating into two streams and further which flows to the atmospheric furnaces F-100 A/B.

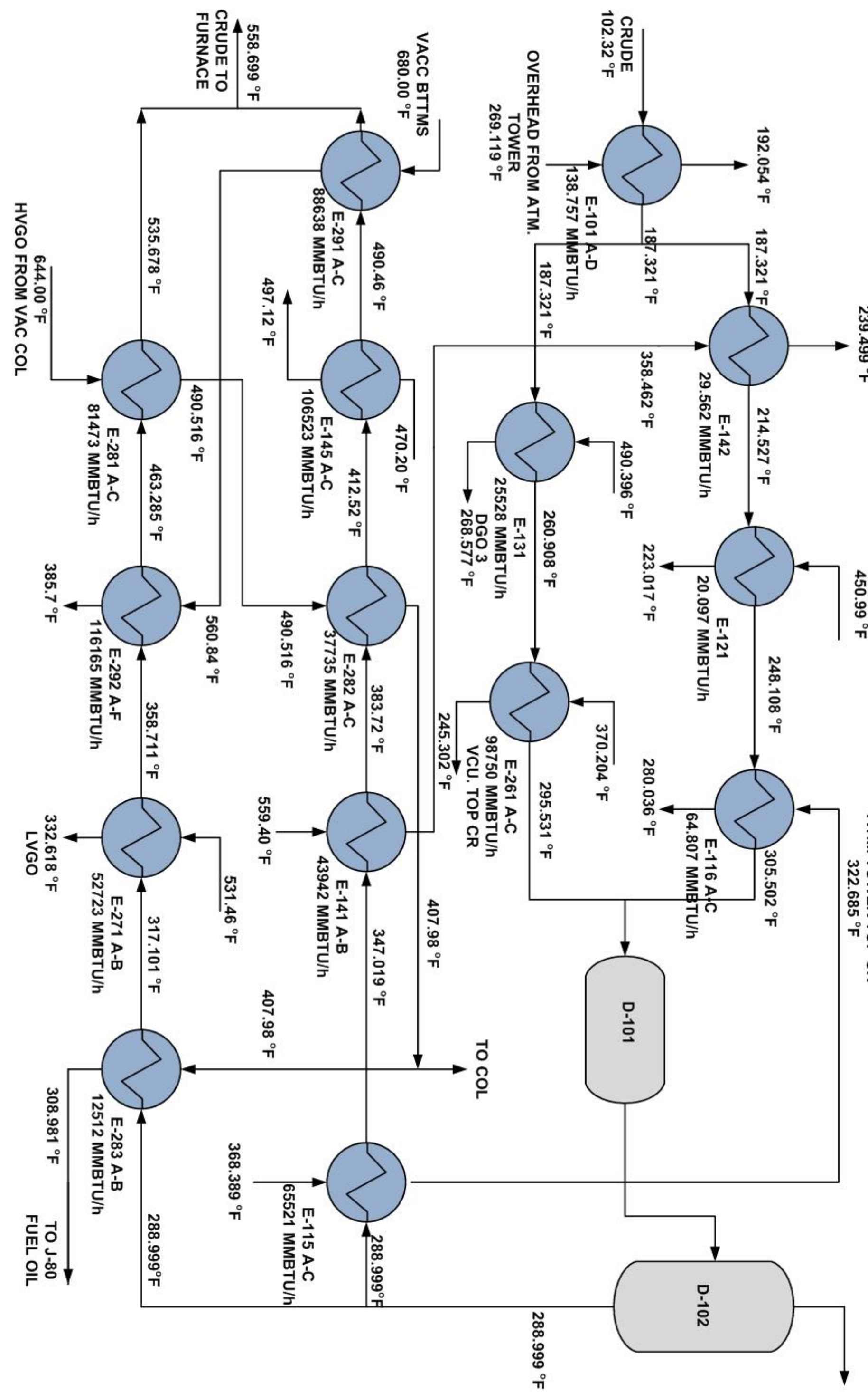


Figure 3.3: Existing Plant Crude Pre-heated Train

3.3 Thermal Data Establishment

In any the process, thermal streams data required for pinch analysis are the heat capacity flow rate (CP), the supply temperature (T_s), the target temperature (T_t) and heat transfer coefficient (h) of hot and cold streams. All information like heat capacity flow rates and temperatures are taken from the process flow sheet. The enthalpy change (ΔH) for the different streams is identified to estimate the heat capacity flowrate and to account for variation in CP with temperature range. The calculation of heat transfer coefficient requires fluid data of stream such as density, velocity, viscosity and geometry of heat transfer surface. In addition, the heat transfer coefficient should include the fouling factor.

3.3.1 Stream Data Extraction

The data are extracted from the crude plant in KSA. It involves identifying relevant hot and cold utility streams from the process flow sheet diagram. The data used in this investigation were those extracted from the refinery between June-2004 and December-2004. The streams data are extracted in four sections: atmospheric section, vacuum section, naphtha section and asphalt section.

3.3.1.1 Atmospheric Section

It is the first section of the plant to separate the crude oil into fractions. It includes primary and secondary crude pre-heat train, flash dram, de-salter and atmospheric distillation. Figures 3.4, 3.5 and 3.6 show a simplified flow diagram of this section and extracted data is given in the Table 3.1. Details of extracted streams are as follows:

- The crude oil exchanges the heat with the overheat stream from the atmospheric column(C-100).
- The crude oil exchanges the heat with the diesel which is side cut three products from the atmospheric column(C-100).
- The crude oil exchanges the heat with the diesel which is side cut four products from the atmospheric column(C-100).
- The crude oil exchanges the heat with the diesel which is side cut tow products from the atmospheric column(C-100).
- The crude oil exchanges the heat with the atmospheric top circulating reflux.
- The crude oil exchanges the heat with the atmospheric top circulating reflux.
- The crude oil exchanges the heat with the diesel which is side cut four products from the atmospheric column(C-100).
- The crude oil exchanges the heat with the atmospheric bottom circulating reflux (C-100).

Table 3.1: Extracted Stream Data for Atmospheric Section

No.	Stream Name	Supply Temperature Ts (^{0}F)	Target Temperature Tt (^{0}F)	Duty (MMBtu/hr)	Flowrate heat capacity CP (MMBtu/hr.F)
1	Diesel	450.99	136	45.542	0.144582
2	Diesel	490.396	140	56.817	0.162151
3	Diesel	570.233	138	97.403	0.225348
4	Diesel	582.483	407.77	177.427	1.015534
5	Kerosene	400	252.459	162.561	1.101802
6	Atoms. OVHD	269.119	192.054	138.757	1.800519
7	Crude Feed	116	187.321	138.757	1.945528
8	Crude Feed	187.321	295.351	134.278	1.24297
9	Crude Feed	187.321	305.502	124.466	1.053181

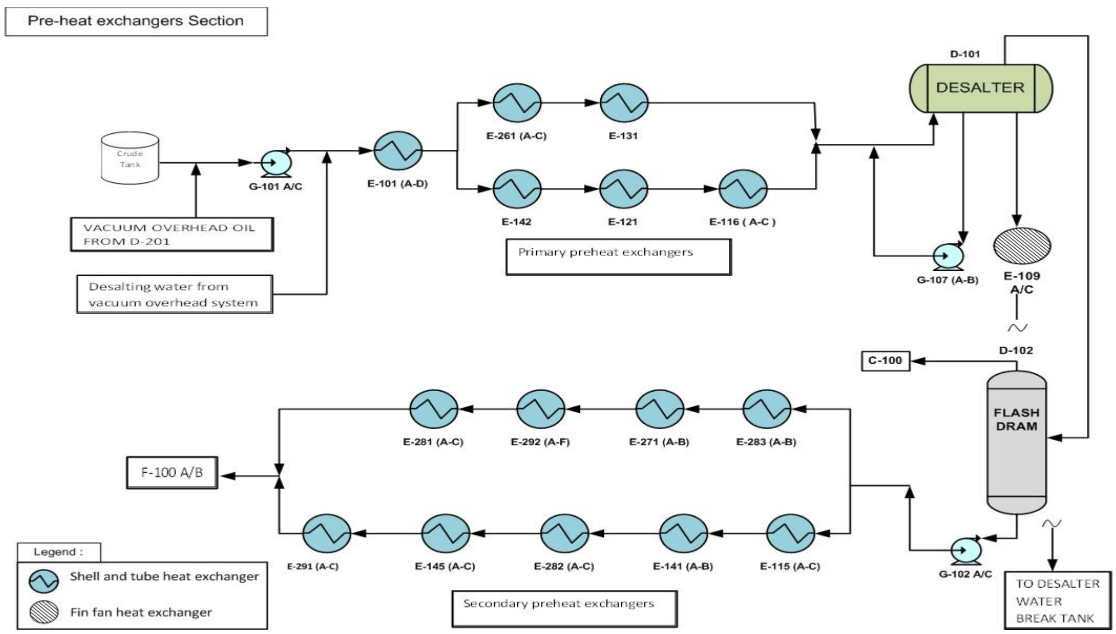


Figure 3.4: Pre-heat Exchangers Section

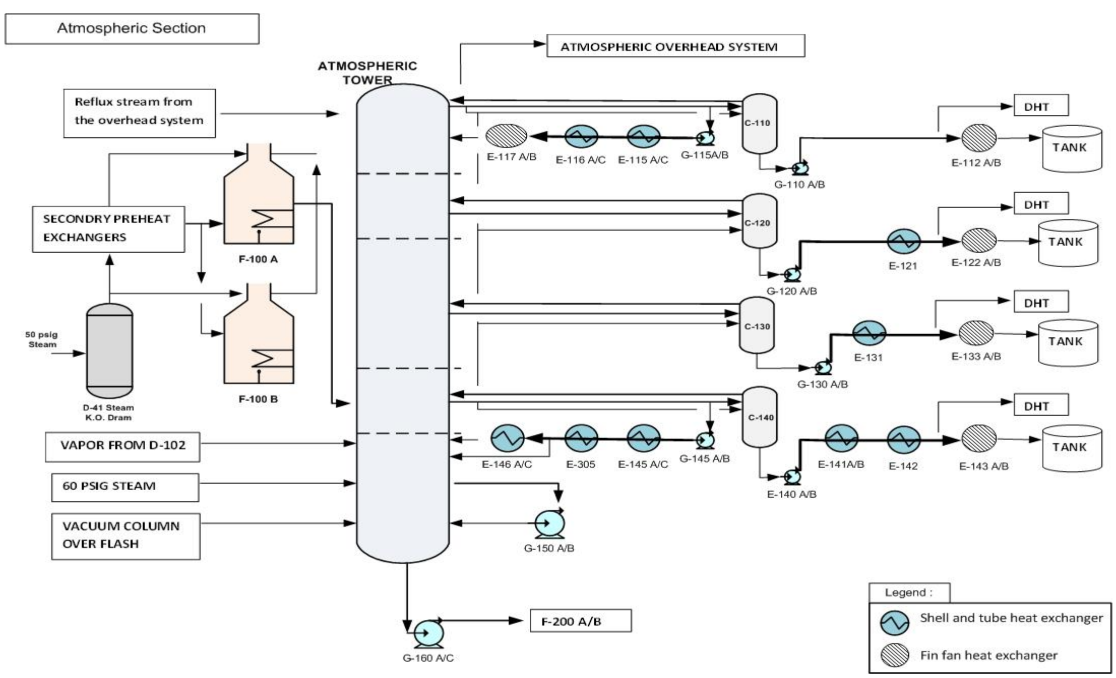


Figure 3.5: Heat Exchangers in Atmospheric Distillation Section

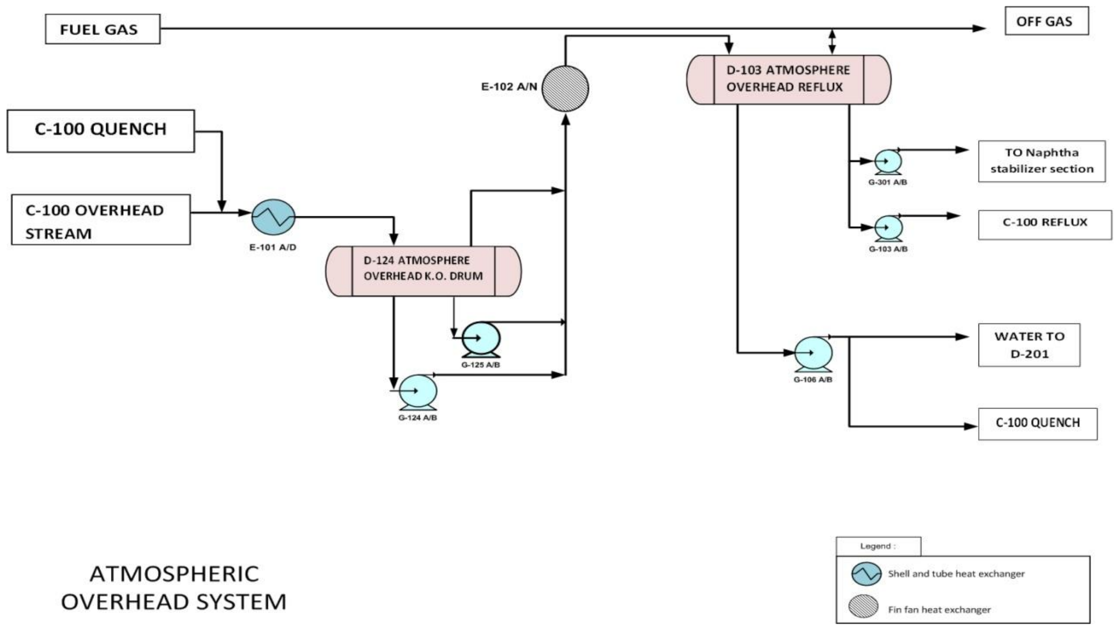


Figure 3.6: Heat Exchangers in Atmospheric Overhead Section

3.3.1.2 Vacuum Section

The vacuum distillation is the main part in this section of the plant that separates the bottom atmospheric product into vacuum gas oil, light vacuum gas oil, heavy vacuum gas oil and vacuum residuum. Figures 3.7 and 3.8 show a simplified flow diagram of this section and extracted data is given in the Table 3.2. Details of extracted streams are as follows:

- The crude oil exchanges the heat with the vacuum gas oil which is side cut six products from the vacuum column(C-200).

- The crude oil exchanges the heat with the light vacuum gas oil which is side cut seven products from the vacuum column(C-200).
- The crude oil exchanges the heat with the heavy vacuum gas oil which is side cut eight products from the vacuum column(C-200).
- The crude oil exchanges the heat with the vacuum residuum which is the vacuum bottom product from the vacuum column(C-200).
- The vacuum overhead stream exchanges the heat with the sea water.

Table 3.2: Extracted Stream Data for Vacuum Section

Symbol	Stream Name	Supply Temperature Ts (^{0}F)	Target Temperature Tt (^{0}F)	Duty (MMBtu/hr)	Flowrate heat capacity CP (MMBtu/hr.F)
1	VGO	370.204	169	136.66	0.679211
2	VGO	245.302	136	7.458	0.068233
3	LVGO	531.46	137.88	92.618	0.235322
4	HVGO	644	394.52	119.208	0.477826
5	HVGO	407.77	308.981	12.512	0.126654
6	Vacuum residuum	681.72	385.7	204.803	0.691855
7	C-200 OVHD	187.902	136	24.03	0.462988
8	C-200 OVHD	199.94	136	30.755	0.480998
9	C-200 OVHD	210.02	137	35.12	0.480964
10	C-200 OVHD	196.5	177	9.4	0.482051
11	Crude Feed	187.321	295.351	134.278	1.24297
12	Crude Feed	187.321	305.502	124.466	1.053181
13	Crude Feed	294.8	669.934	483.313	1.288374
14	Crude Feed	294.8	669.96	581.687	1.550504

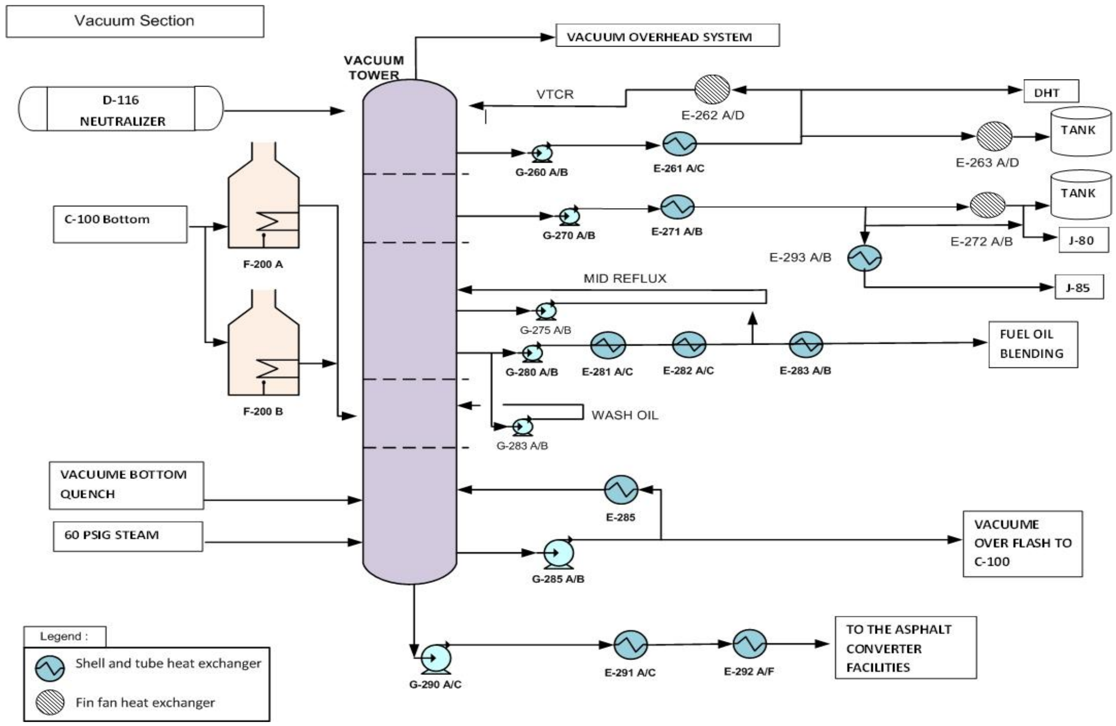


Figure 3.7: Heat Exchangers in Vacuum Distillation Section

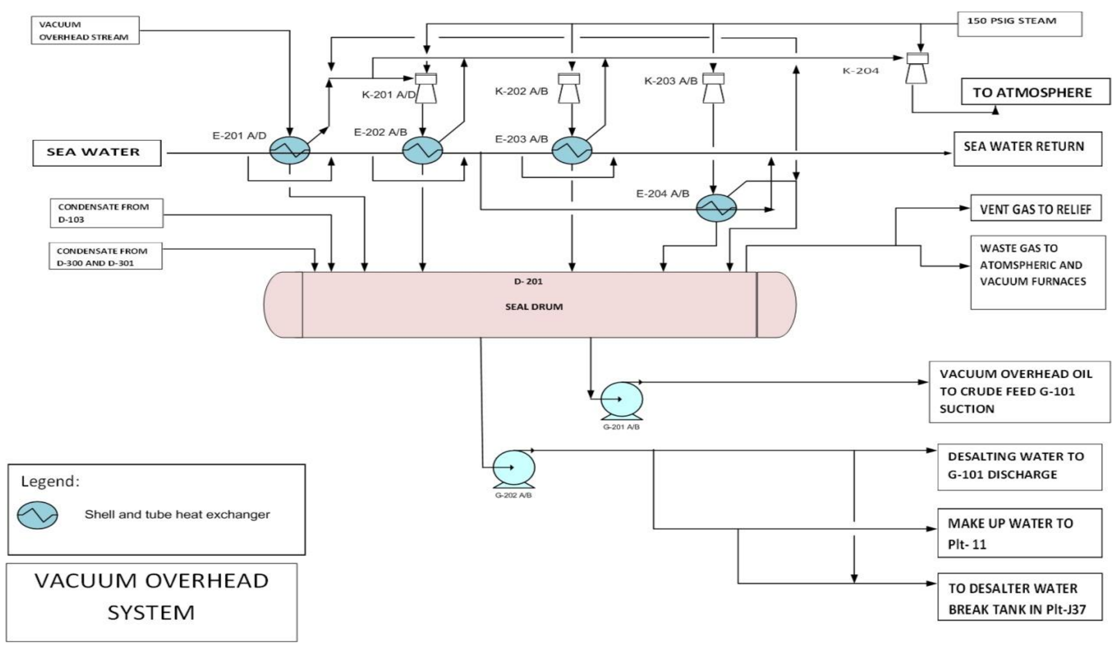


Figure 3.8: Heat Exchangers in Vacuum Overhead Section

3.3.1.3 Naphtha Section

The naphtha column is the main part in this section of the plant that separates the stabilized naphtha from the LPG. Figure 3.9 show a simplified flow diagram of this section and extracted data is given in the Table 3.3. Detail of extracted streams is as follow:

- The naphtha feed exchanges the heat with the naphtha stabilizer column product stream from the naphtha column(C-300).
- The diesel exchanges the heat with the naphtha stream from the naphtha column(C-300).

Table 3.3: Extracted Stream Data for Naphtha Section

NO.	Stream Name	Supply Temperature Ts (^{0}F)	Target Temperature Tt (^{0}F)	Duty (MMBtu/hr)	Flowrate heat capacity CP (MMBtu/hr.F)
1	Diesel	582.483	407.77	177.427	1.015534
2	Naphtha	359.606	203.809	80.116	0.514233
3	LPG	175.208	136	7.812	0.199245
4	Naphtha	179.218	136	31.0155	0.717652
5	Naphtha feed	123.862	311.263	80.116	0.427511
6	Naphtha	337.767	370.173	70.90399	2.18799

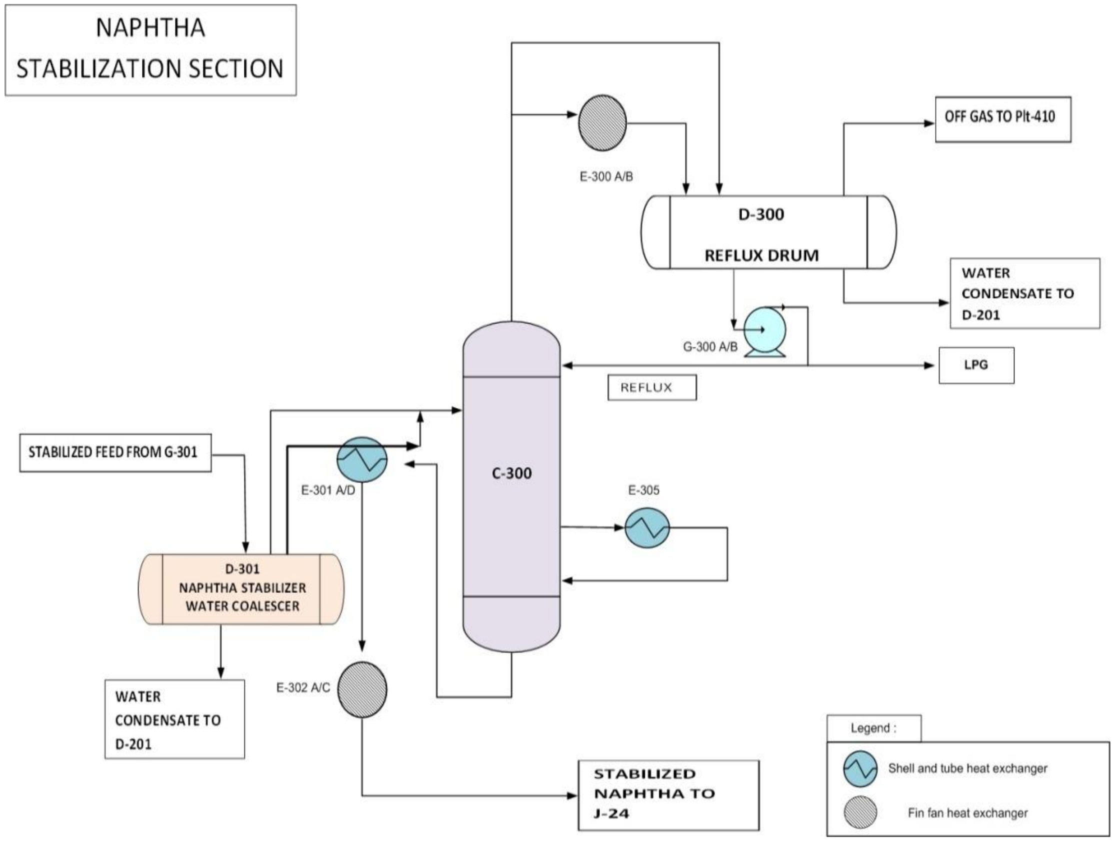


Figure 3.9: Heat Exchangers in Naphtha Stabilization Section

3.3.1.4 Asphalt Section

In this section of the plant receives the vacuum residuum from the vacuum section, the feed exchanges with the boiler feed water. Figure 3.10 show a simplified flow diagram of this section and extracted data is given in the Table 3.4. Details of extracted streams are as follows:

- The vacuum residuum exchanges the heat with the boiler feed water stream.
- The paving asphalt exchanges the heat with the boiler feed water stream.

Table 3.4: Extracted Stream Data for Asphalt Section

NO.	Stream Name	Supply Temperature Ts (^{0}F)	Target Temperature Tt (^{0}F)	Duty (MMBtu/hr)	Flowrate heat capacity CP (MMBtu/hr.F)
1	Vacuum residuum	407.084	326.034	24.01	0.296237
2	Paving asphalt	442.858	376.985	57.086	0.866607
3	Paving asphalt	490.396	368.577	10.881	0.089321

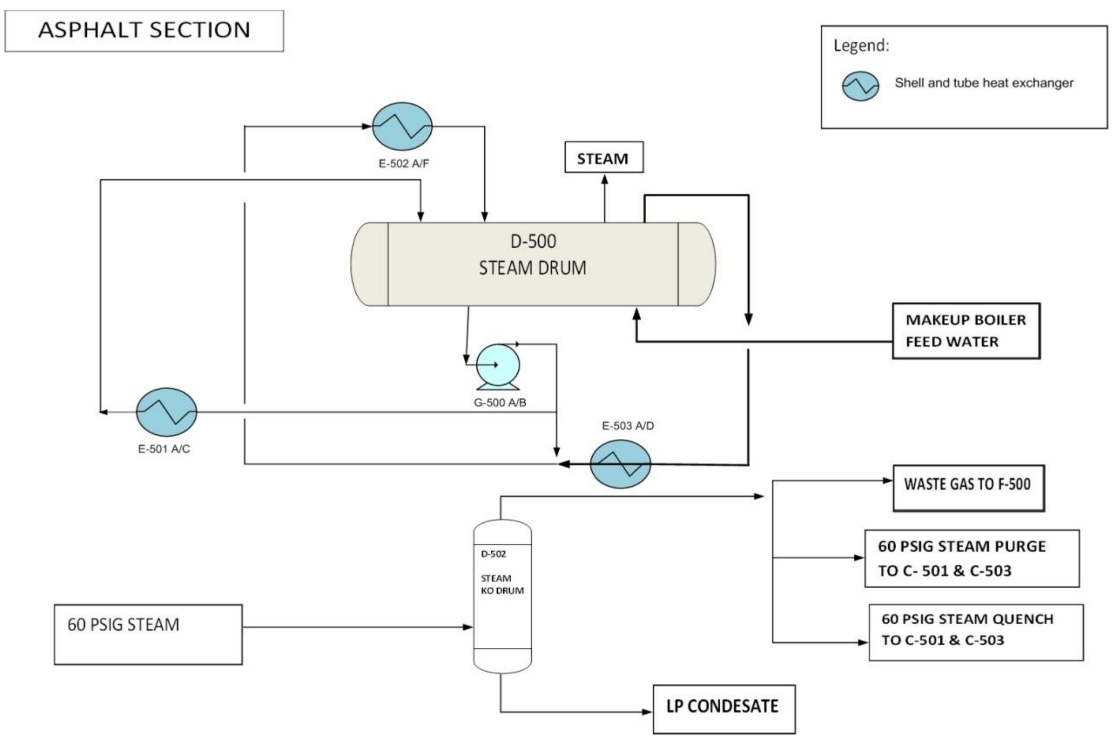


Figure 3.10: Heat Exchangers in Asphalt Section

3.4 Estimation of Cost Data

Economic assessment of energy saving opportunity is of paramount importance in retrofitting analysis. The cost is annualized to study the yearly savings accruable from the retrofits in economic terms and to estimate the payback period. A useful life period of 12 years and an annual interest rate of 6% were adopted. It was assumed that plant operates for 7920 hours per year (Al-Riyami, et al., 2001). The capital cost estimation is based on the heat exchanger unit costs.

3.4.1 Energy Cost

The furnaces are fired using fuel oil and flue gases from the four furnaces are available for use as part of hot utility sources the cold utilities includes air-cooling, water-cooling and boiler feed water. The energy cost for different $\Delta T_{\min}$ is calculated using the Equation:

$$Energy\ Cost\ =\ \sum Q_u\ \times\ C_u \tag{3.1}$$

Where Q_u is the duty of utility U, MMBtu; C_u is the unit cost of utility U, \$/MMBtu yr and is $\sum$ the cost summation for all utilities used. For the utility cost calculation, the cost data presented in Table 3.5 were used.

Table 3.5: Utility Costs

Utility	Unit cost ($/MMBtu)
Fuel gas cost	19.33
Air-cooling	0.472
Water-cooling	0.718
Boiler feed water	5.847

The annual costs of the utilities in used for this investigation as follows;

- Fuel gas cost = 153093.6 $/(MMBtu/hr) yr
- Air-cooling = 3738.24 $/(MMBtu/hr) yr
- Water-cooling = 7180 $/(MMBtu/hr) yr
- Boiler feed water = 46308.24 $/(MMBtu/hr) yr

3.4.2 Capital Cost

The capital annualized capital cost for the heat exchanger for different $\Delta T_{\min}$ is calculated using the following Equation (2.2) (Al-Riyami et al., 2001):

$$Heat\ Exchanger\ Cost\ (HEC) = A + B\ (Area)^c \tag{2.2}$$

Where A represents a fixed cost of installation independent of area, B the exchanger cost per unit area and which also accounts for different materials of construction. The type of heat exchanger used in the analysis is the stainless still shell and tube heat exchanger (ss). The heat exchanger cost for this type of heat exchanger is calculated using Equation (2.3):

$$Heat\ Exchanger\ Cost\ (HEC) = 33422 + 1784\ (Area)^{0.61} \tag{2.3}$$

The total cost is the summation of capital and energy cost:

$$Total\ Cost\ = Energy + Capital\ Cost \tag{2.4}$$

3.5 Discussion

The crude distillation process heat exchanger network has been described. The scope for energy saving of the existing network can be identified by applying pinch technology analysis tools to measure the performance and set targets for the energy recovery. Collecting and extracting data from the process is first step of the pinch analysis and it is important to extract essential features of the process such as the supply and target temperatures and enthalpy changes of the streams. The thermal data required for the analysis also needs the specification of the heat transfer coefficients which should include the fouling factors.

Stream fluid properties and geometry of exchanger gives information about calculation of heat transfer coefficients. The existing heat exchangers network of the plant consists of twenty-eight hot and nine cold streams given in the form required for pinch analysis, in the Table 3.6.

Table 3.6: Thermal Data of Crude Plant Used for the Pinch Analysis of the Process

Symbol	Stream Name	Supply Temperature Ts (^{0}F)	Target Temperature Tt (^{0}F)	Duty (MMBtu/hr)	Flowrate heat capacity CP (MMBtu/hr.F)
HS-1	Diesel	450.99	136	45.542	0.144582
HS-2	Diesel	490.396	140	56.817	0.162151
HS-3	Diesel	570.233	138	97.403	0.225348
HS-4	VGO	370.204	169	136.66	0.679211
HS-5	VGO	245.302	136	7.458	0.068233
HS-6	LVGO	531.46	137.88	92.618	0.235322
HS-7	HVGO	644	394.52	119.208	0.477826
HS-8	HVGO	407.77	308.981	12.512	0.126654
HS-9	Diesel	582.483	407.77	177.427	1.015534
HS-10	Kerosene	400	252.459	162.561	1.101802
HS-11	Vacuum residuum	681.72	385.7	204.803	0.691855
HS-12	Atoms. OVHD	269.119	192.054	138.757	1.800519
HS-13	Deiesel	445.585	420.182	40.034	1.575956
HS-14	vac. Overflash	674.04	454.885	5.601	0.025557
HS-15	Vacuum residuum	416.794	349.469	27.468	0.407991
HS-16	Vacuum residuum	407.084	326.034	24.01	0.296237
HS-17	Paving asphalt	442.858	376.985	57.086	0.866607
HS-18	Paving asphalt	490.396	368.577	10.881	0.089321
HS-19	Naphtha	359.606	203.809	80.116	0.514233
HS-20	LPG	175.208	136	7.812	0.199245
HS-21	Naphtha	179.218	136	31.0155	0.717652
HS-22	Atmos. OVHD	193.552	137	2.85	0.050396
HS-23	Desalter Effluent water	309.92	138	11.495	0.066862
HS-24	Kerosene	335.689	138	25.301	0.127984
HS-25	C-200 OVHD	187.902	136	24.03	0.462988
HS-26	C-200 OVHD	199.94	136	30.755	0.480998
HS-27	C-200 OVHD	210.02	137	35.12	0.480964
HS-28	C-200 OVHD	196.5	177	9.4	0.482051
CS-1	Crude Feed	116	187.321	138.757	1.945528
CS-2	Crude Feed	187.321	295.351	134.278	1.24297
CS-3	Crude Feed	187.321	305.502	124.466	1.053181
CS-4	Crude Feed	294.8	669.934	483.313	1.288374
CS-5	Crude Feed	294.8	669.96	581.687	1.550504
CS-6	Naphtha feed	123.862	311.263	80.116	0.427511
CS-7	Naphtha	337.767	370.173	70.90399	2.18799
CS-8	Crude Feed	618.043	749.055	113.216	0.864165
CS-9	Crude Feed	618.043	749.927	107.246	0.813184

CHAPTER 4

PINCH ANALYSIS TARGETING

4.1 Introduction

The extraction of the data is the first step for pinch analysis. They consist of thermodynamic and economic data. The composite curve is used for minimum energy target of the process and based on thermodynamic principles. The minimum network approach temperature needs to be defined for composite curve. This is defined by the aim of project and the economics of the process.

The objective of the project is to reduce the energy costs. The network is designed to achieve the target. The target of retrofit design is different from a grassroot design. In the retrofit case, the design of the optimum heat exchanger network is constrained by the existing design of the network. Furthermore, the economics of the process determines the optimum heat recovery. The target for the design of heat exchanger network of plant is set and the existing network in analyzed. The network pinch method is applied of the design of heat exchanger network. The results are compared with the targets.

4.2 Existing Heat Exchanger Network

The process data consists of twenty-eight hot and nine cold streams. These were given in in the Table 3.1 in the previous chapter. Furthermore, Figure 4.1 shows the existing grid diagram of the process which has seventeen heat exchangers process-to-process in the Table 4.1, twenty-two cold utility units that include four water coolers, twelve air coolers and six boiler feed water coolers in the Tables 4.2, 4.3 and 4.4 respectively, and four hot utility units that use four furnaces in the Table 4.5.

The heat load of existing heat exchangers network is found from plotting composite curve of hot and cold streams. The existing energy consumption predicting are as follows:

Hot utility consumption = 680.23 MMBtu/hr, Cold utility consumption = 521.215 MMBtu/hr, All utilities consumption correspond to ΔT_{min} = 77 ^{0}F.

Table 4.1: Process-to-Process Heat Exchangers Data

HE. NO#	Exchanger Name	Area (ft^2)	Duty (MMBtu/hr)	Hot Stream		Flowrate heat capacity CP (MMBtu/hr.F)	Cold Stream		Flowrate heat capacity CP (MMBtu/hr.F)
				T_s (^{0}F)	T_t (^{0}F)		T_s (^{0}F)	T_t (^{0}F)	
1	E-101(A-D)	31743	138.757	269.119	192.054	1.8005	116	187.321	1.94552797
2	E-115(A-C)	19394	65.521	368.389	322.685	1.433594	294.8	347.019	1.129283
3	E-116(A-C)	19394	64.807	322.685	280.036	1.51954	248.108	305.502	1.129159
4	E-121	4309	30.097	450.99	223.017	0.13202	214.527	248.108	0.89625
5	E-131	4400	35.528	490.396	268.577	0.160166	187.321	260.908	0.482802
6	E-141(A-B)	5584	43.942	570.233	358.462	0.207497	347.019	386.081	1.12492
7	E-142	4398	29.562	358.462	239.499	0.24849	187.321	214.527	1.086598
8	E-145(A-B)	21006	106.523	582.482	478.791	1.02731	421.465	529.454	0.986424
9	E-261(A-C)	23986	98.75	370.204	245.302	0.790619	260.908	295.351	2.867055
10	E-271(A-B)	5591	52.723	531.46	332.618	0.26515	317.101	358.711	1.267072
11	E-281(A-C)	21006	81.473	644	490.516	0.530824	463.285	535.678	1.125426
12	E-282(A-C)	19394	37.735	490.516	407.98	0.457194	386.081	421.465	1.066442
13	E-283(A-B)	10743	12.512	407.98	308.981	0.126385	294.8	317.101	0.56105107
14	E-291(A-D)	20941	88.638	681.72	560.84	0.73327	490.45.454	551.66	1.695901
15	E-292(A-F)	41882	116.165	560.84	385.7	0.663269	358.711	463.285	1.11084
16	E-301(A-D)	41882	80.116	359.606	203.809	0.514233	357.98	461.12	0.776769
17	E-305	21006	70.904	478.791	407.77	0.998352	337.767	370.173	2.187989

Table 4.2: Cold Utility Units, Using cooling Water, Heat Exchanger Data

HE. NO#	Exchanger Name	Area (ft^2)	Duty (MMBtu/hr)	Hot Stream		Flowrate heat capacity CP (MMBtu/hr.F)	Cold Stream		Flowrate heat capacity CP (MMBtu/hr.F)
				T_s (^0F)	T_t (^0F)		T_s (^0F)	T_t (^0F)	
18 cu	E-201 (A-D)	5425	24.03	187.902	136	0.462987939	59	86.466	0.87489988
19 cu	E-202 (A-B)	5823	30.755	199.94	136	0.48099781	87.847	96.854	3.41456645
20 cu	E-203 (A-B)	3175	35.12	210.02	137	0.480964119	97.798	103.706	5.94448206
21 cu	E-204 (A-B)	1808	9.4	196.5	177	0.482051282	97.798	109.447	0.80693622

Table 4.3: Cold Utility Units, Using Air Cooling, Heat Exchanger Data

HE. NO#	Exchanger Name	Area (ft^2)	Duty (MMBtu/hr)	Hot Stream		Flowrate heat Capacity CP (MMBtu/hr.F)	Cold Stream		Flowrate heat Capacity CP (MMBtu/hr.F)
				T_s (^0F)	T_t (^0F)		T_s (^0F)	T_t (^0F)	
22 cu	E-102(A-N)	1363	2.85	193.552	121.498	0.0395537	193.552	121.498	0.0395537
23 cu	E-109 (A-C)	3576	11.495	309.92	123.516	0.0616671	309.92	123.516	0.0616671
24 cu	E-112 (A-B)	9731	25.301	335.689	123.251	0.3827046	335.689	123.251	0.3827046
25 cu	E-117 (A-B)	4063	32.233	280.036	252.459	1.1688363	280.036	252.459	1.1688363
26 cu	E-122(A-B)	2828	15.445	223.017	123.461	0.1551388	223.017	123.461	0.1551388
27 cu	E-133 (A-B)	26167	21.289	268.577	111.433	0.1354745	268.577	111.433	0.1354745
28 cu	E-143(A-B)	4084	23.899	239.499	115.067	0.1920647	239.499	115.067	0.1920647
29 cu	E-262(A-D)	3204	37.91	245.302	186.752	0.6474808	245.302	186.752	0.6474808
30 cu	E-263(A-B)	3204	7.458	245.302	113.143	0.056432	245.302	113.143	0.056432
31 cu	E-272(A-B)	6827	39.895	332.618	137.88	0.204865	332.618	137.88	0.204865
32 cu	E-300 (A-B)	2576	7.812	175.208	103.328	0.1086811	175.208	103.328	0.1086811
33 cu	E-302(A-C)	5529	31.0155	179.218	120.27	0.5261502	179.218	120.27	0.5261502

Table 4.4: Cold Utility Units, Using Boiler Feed Water, Heat Exchanger Data

HE. NO#	Exchanger Name	Area (ft^2)	Duty (MMBtu/hr)	Hot Stream		Flowrate heat capacity CP (MMBtu/hr.F)	Cold Stream		Flowrate heat capacity CP (MMBtu/hr.F)
				T_s (^{0}F)	T_t(^{0}F)		T_s (^{0}F)	T_t(^{0}F)	
34 cu	E-146	2634	40.034	445.585	420.182	1.575955596	367.148	372	8.2510305
35 cu	E-285	764	5.601	674.04	454.885	0.025557254	367	369	2.8005
36 cu	E-293A/D	7893	27.468	416.794	349.469	0.407991088	308.826	354.625	0.59975109
37 cu	E-501A/C	1294	24.229	407.084	326.034	0.298938927	316.707	318.569	13.0123523
38 cu	E-502A/F	1294	57.086	442.858	376.985	0.866606956	316.707	321.677	11.4861167
39 cu	E-503A/D	300	10.881	490.396	268.577	0.049053508	316.707	319	4.74531182

Table 4.5: Hot Utility Units, Heat Exchanger Data

HE. NO#	Exchanger Name	Type	Duty (MMBtu/hr)	T_s(^{0}F)	T_t (^{0}F)	Flowrate heat capacity (MMBtu/hr.F)	T_s (^{0}F)	T_t (^{0}F)	Flowrate heat capacity (MMBtu/hr.F)
40 hu	F-100A	Hot	220.44	1800	626	0.18776831	498.834	669.934	1.288369
41 hu	F-100B	Hot	239.328	1800	632	0.20490411	515.6	669.96	1.550453
42 hu	F-200A	Hot	113.216	1800	621	0.09602714	618.043	749.055	0.864165
43 hu	F-200B	Hot	107.246	1800	621	0.09096353	618.043	749.055	0.818597

4.3 Establishment of ΔTmin and Energy Targeting

The limit of the heat recovery of process is given by the composite curve. The minimum approach temperature is required to define the composite curve. The setting of the minimum approach temperature is economic. There is a trade-off between energy and capital cost. For grassroot designs, the total annualized cost is plotted versus ΔT_{min} for a range of energy recovery. From this plot, the optimum value of the ΔT_{min}, which gives the lowest total cost is found. However for retrofit, we already have an existing design with its existing recovery and area. To improve the energy recovery, we often need to invest capital for more exchangeing area. We need to trade off investment with energy savings. This is either limited by investment or payback period of project.

To tradeoff between energy and capital, costs have to be established in the same basis. Capital costs need to be annualized as well as the economic data of the process have to be specified such as plant life, interest rate and annual operating hours. The data of utility and economic used in the pinch analysis of process are given in the table 4.6 and 4.7 respectively.

To improve heat recovery, we need to reduce minimum approach temperature (ΔT_{min}) and trade-off between energy and area to obtain an optimal $\Delta T_{min.}$

Table 4.6: Utility Consumption Data

Stream Name	Type	Duty (MMBtu/hr)	Cost $/(MMBtu/hr)yr
Flue gas	Hot	680.23	153093.6
Boiler Feed Water	Cold	165.308	46308.24
Cooling Air	Cold	256.6025	3738.24
Cooling Water	Cold	99.305	7180

Table 4.7: Economic Data

Constants	**Values**
Annualized Factor Function	$\frac{i(i+1)^n}{(i+1)^n - 1}$
Interest rate	6 %
Plant life	12 years
Annualization factor	0.1193
Annual operating time	7920 hrs

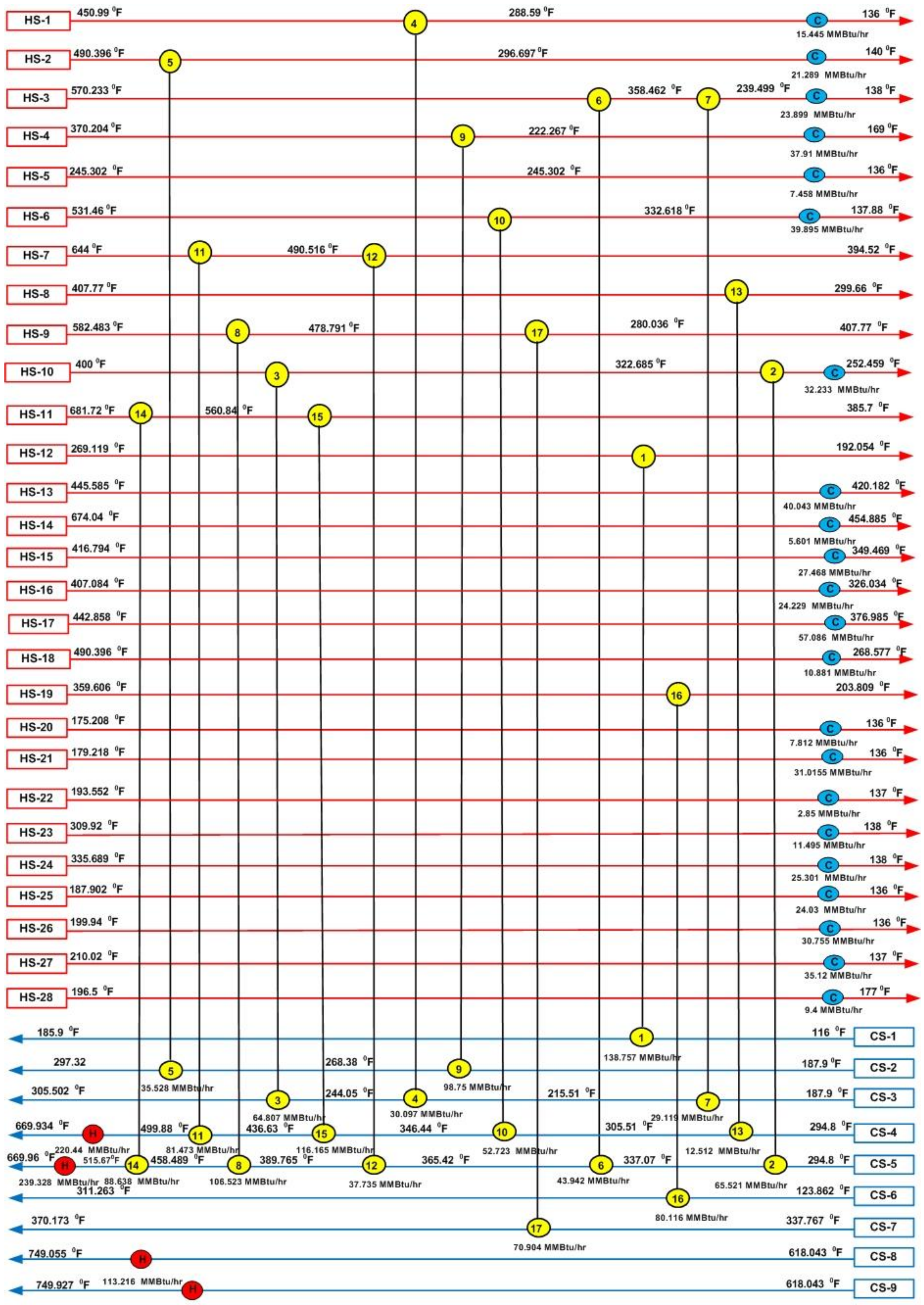


Figure 4.1: The Existing Grid Diagram of the Process

4.3.1 Total Annualized Cost Plot

The costs of heating and cooling utilities and cost of heat exchangers evaluation were conducted at different ΔT_{min} in other to obtain the minimum driving force required for optimum trade-off balance between capital cost and energy cost. Where annualized total cost (A1-1) and (A1-2) against minimum approach temperature are drawn in Figure 4.2 and Figure 4.3 respectively. From these figures, we can be shown that both plots of the total cost decline from a high value until a specific value of ΔT_{min} and then increase again. In addition, we can note when ΔT_{min} has large value, the capital cost is low but the energy cost is high. In contrast, when ΔT_{min} has small value, the capital cost is high but the energy cost is low. Therefore, the optimum value of ΔT_{min} is between them. From cost report these plots in appendix A (Table 4.8 A1-1 and Table 4.9 A1-2), it is obtained that optimum values of ΔT_{min} are 38 ^{0}F and 76 ^{0}F respectively. It is often the A1-2 cost has larger value than A1-1 due to the capital cost of heat exchanger where there is extra area required.

To evaluate the minimum approach temperature from total cost, we choose average between temperatures, (57 ^{0}F), therefore we select the optimum (ΔT_{min}= 57 ^{0}F).

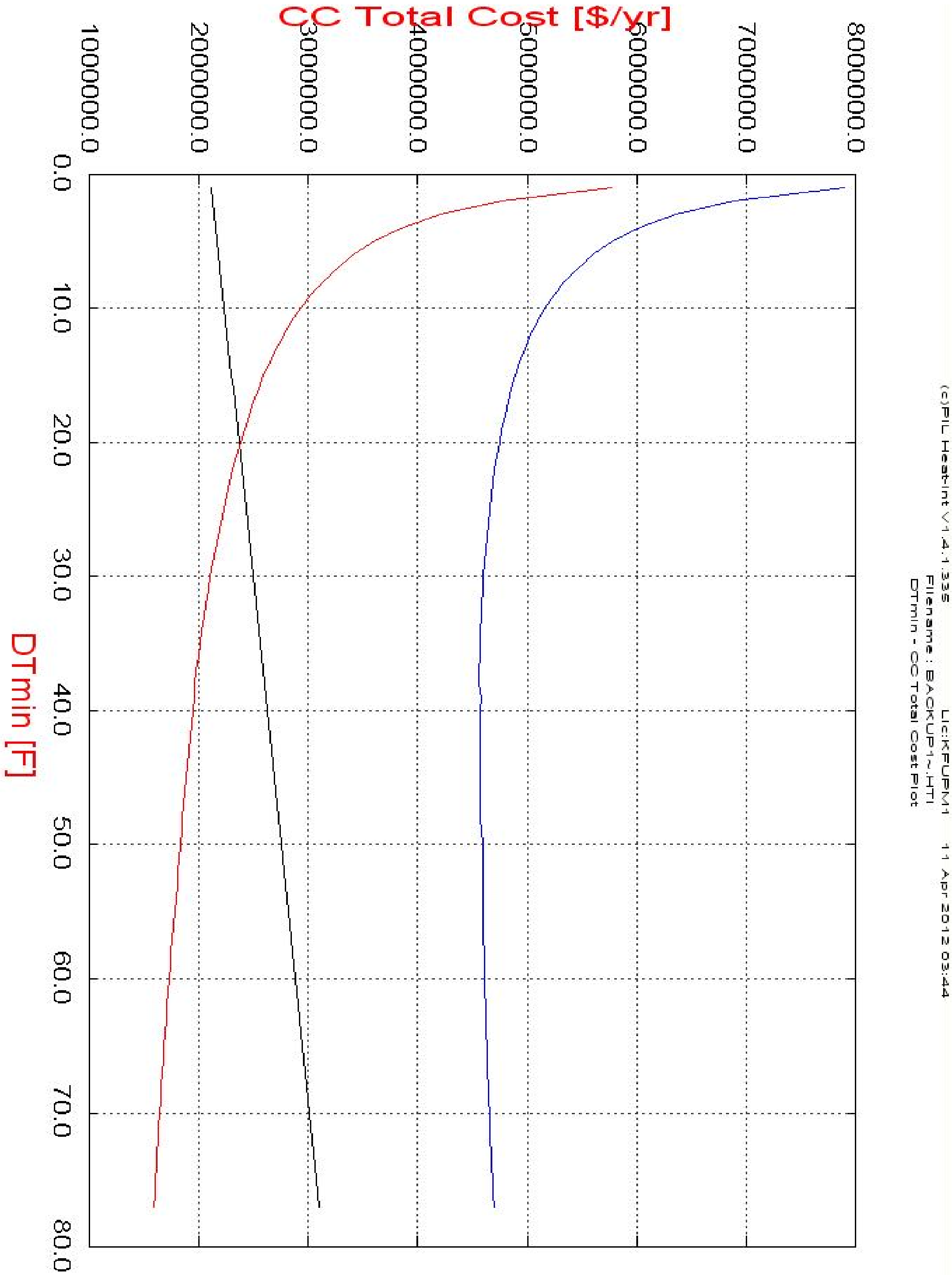


Figure 4.2: Range Target Plot of Total Annualized Cost (A1-1) Showing the Optimum ΔT_{min}

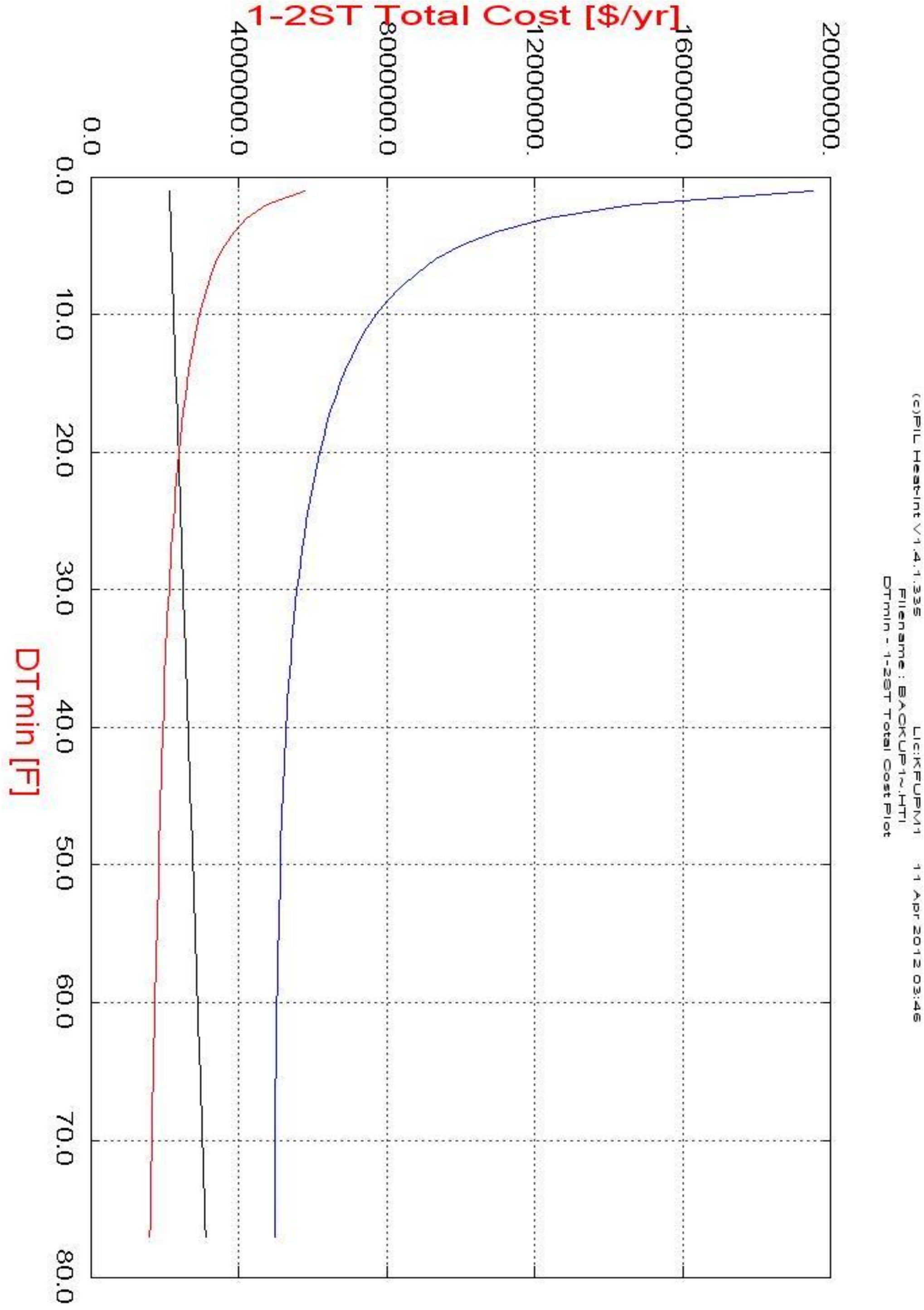


Figure 4.3: Range Target Plot of Total Annualized Cost (A1-2) Showing the Optimum ΔT_{min}

Table 4.8: Values of Range Target Plot of Total (A1-1) Cost Versus in the Region of Optimum Value

$\Delta T_{min}(^0F)$	Total (A1-1) cost $/yr
30.0	0.4601E+07
31.0	0.4593E+07
32.0	0.4586E+07
33.0	0.4580E+07
34.0	0.4575E+07
35.0	0.4570E+07
36.0	0.4566E+07
37.0	0.4563E+07
* 38.0	0.4560E+07 *
39.0	0.4577E+07
40.0	0.4575E+07

Table 4.9: Values of Range Target Plot of Total (A1-2) Cost Versus in the Region of Optimum Value

$\Delta T_{min}(^0F)$	Total (A1-2) cost $/yr
70.0	0.4973E+07
71.0	0.4977E+07
72.0	0.4968E+07
73.0	0.4972E+07
74.0	0.4963E+07
75.0	0.4968E+07
* 76.0	0.4959E+07 *
77.0	0.4964E+07

4.3.2 Determination of ΔTmin for Retrofit Design

Retrofit target technique for existing heat exchangers network based on the minimum approach temperature to achieve the reduction of utility consumption. After we found the minimum approach temperature (ΔT_{min}= 57 ^{0}F), and plotted the composite and grand curves at optimum ΔTmin (Figures 4.4 and 4.5) in order to estimate the location of the pinch, above and below pinch temperature. For these Figures 4.4 and 4.5, the pinch point temperature can be obtained at 461.9 ^{0}F (shifted temperature), corresponding to the above pinch temperature ($T_{above\ pinch}$ = 490.4 ^{0}F) and the below pinch temperature ($T_{below\ pinch}$ = 433.4 ^{0}F).

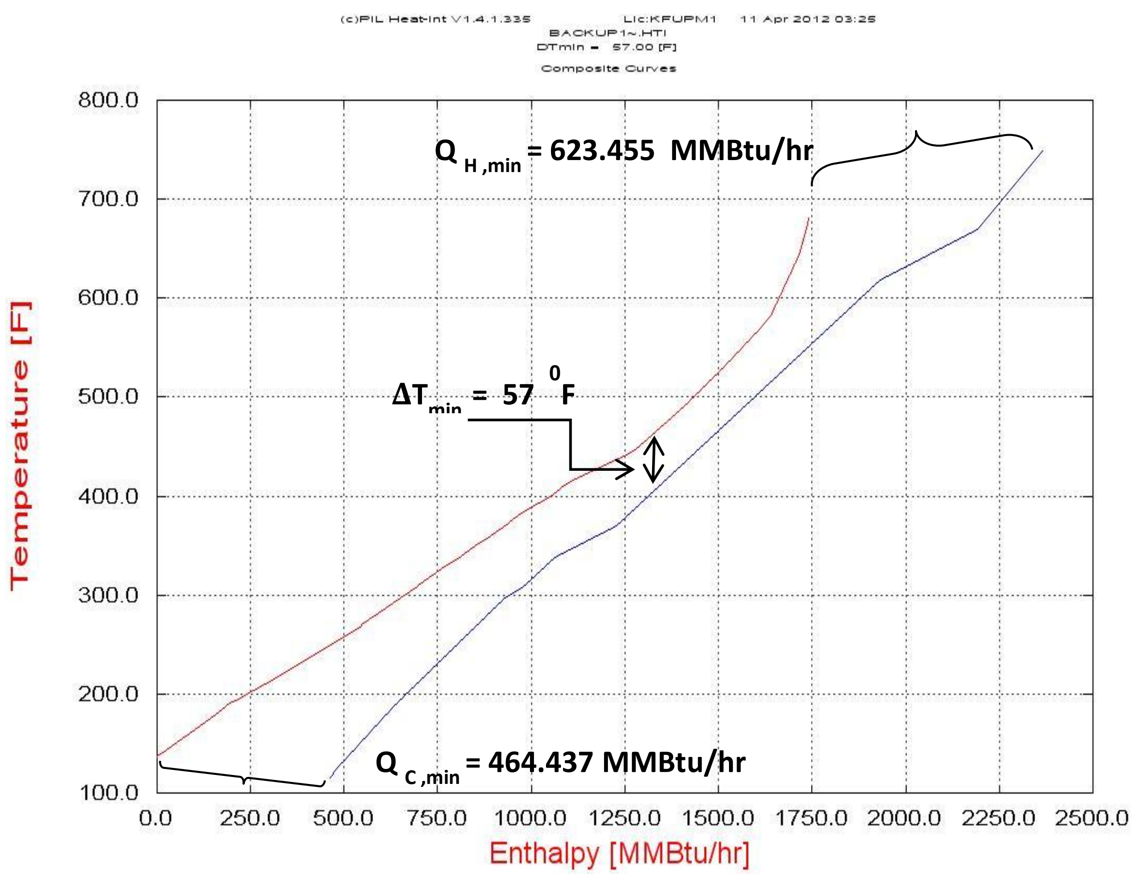


Figure 4.4: Composite Curves of the Process

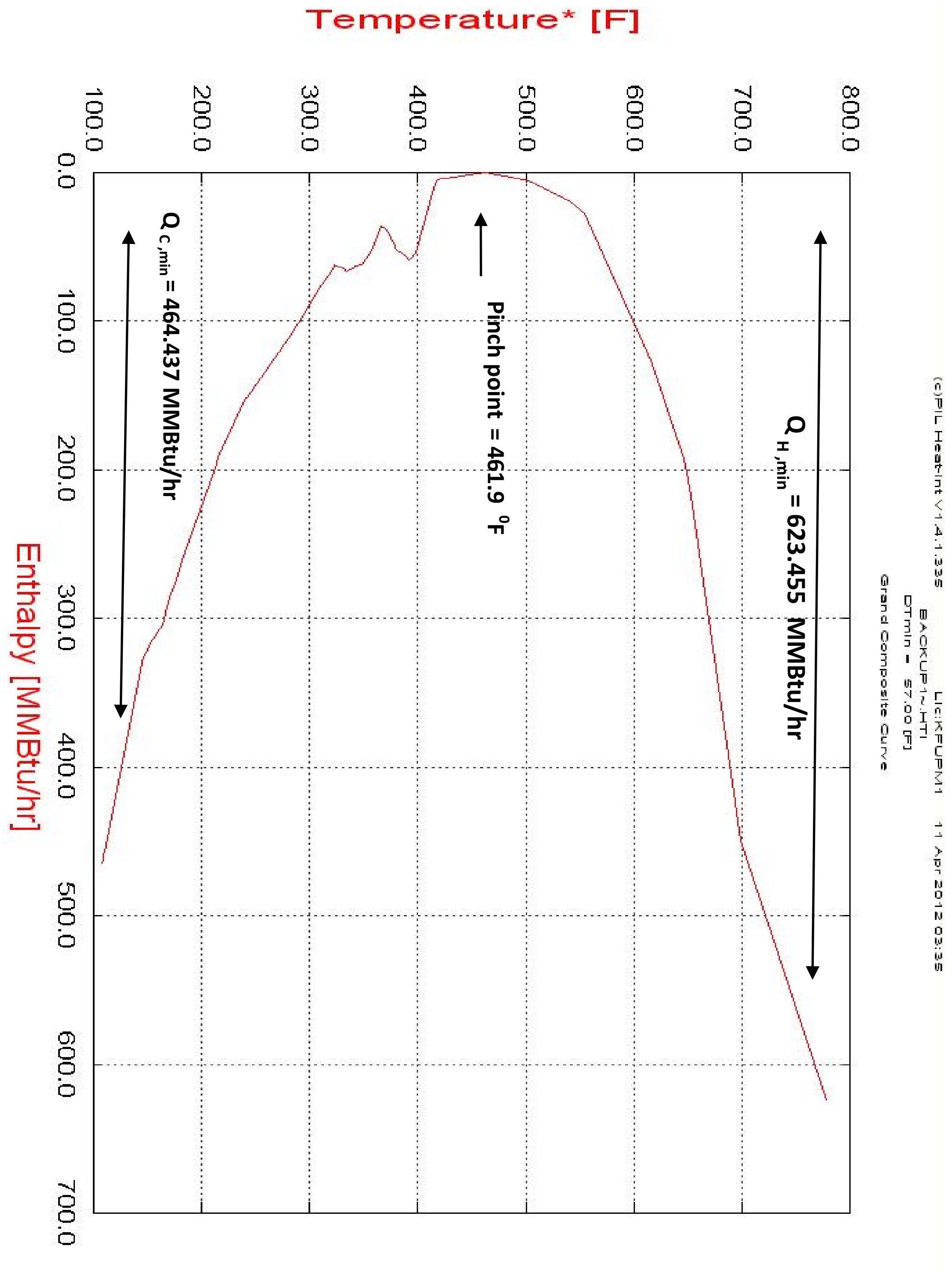


Figure 4.5: Grand Composite Curves of the Process

4.4 Analysis of Existing Network

At the targeting stage for retrofitting of the heat exchangers network found some heat exchangers that violation rules (no heat transfer across pinch, at above the pinch do not use cold utilities and at below the pinch do no use hot utilities).

The heat exchangers that transfer heat across pinch and inappropriate utility placement are identified. Figure 4.6 shows process to process heat exchanger units that transfer heat across the pinch from above to below the pinch in the existing design which include HE-6, HE-8, HE-10, HE-11 and HE-15. Also, Figure 4.7 show cold utility units that are used inappropriately in the existing design removing heat from above the pinch, which include HE-18cu, HE-19cu and HE-35cu. Summary of the heat flow violation of pinch is given in below table 4.10.

Table 4.10: Heat Exchangers that Transfer Heat Across-pinch

HE-6	HE-8	HE-10	HE-15	HE-18cu	HE-19cu	HE-35cu
17.909 (MMBtu/hr)	54.6901 (MMBtu/hr)	9.663 (MMBtu/hr)	40.6485 (MMBtu/hr)	20.7096 (MMBtu/hr)	21.7175 (MMBtu/hr)	4.6934 (MMBtu/hr)

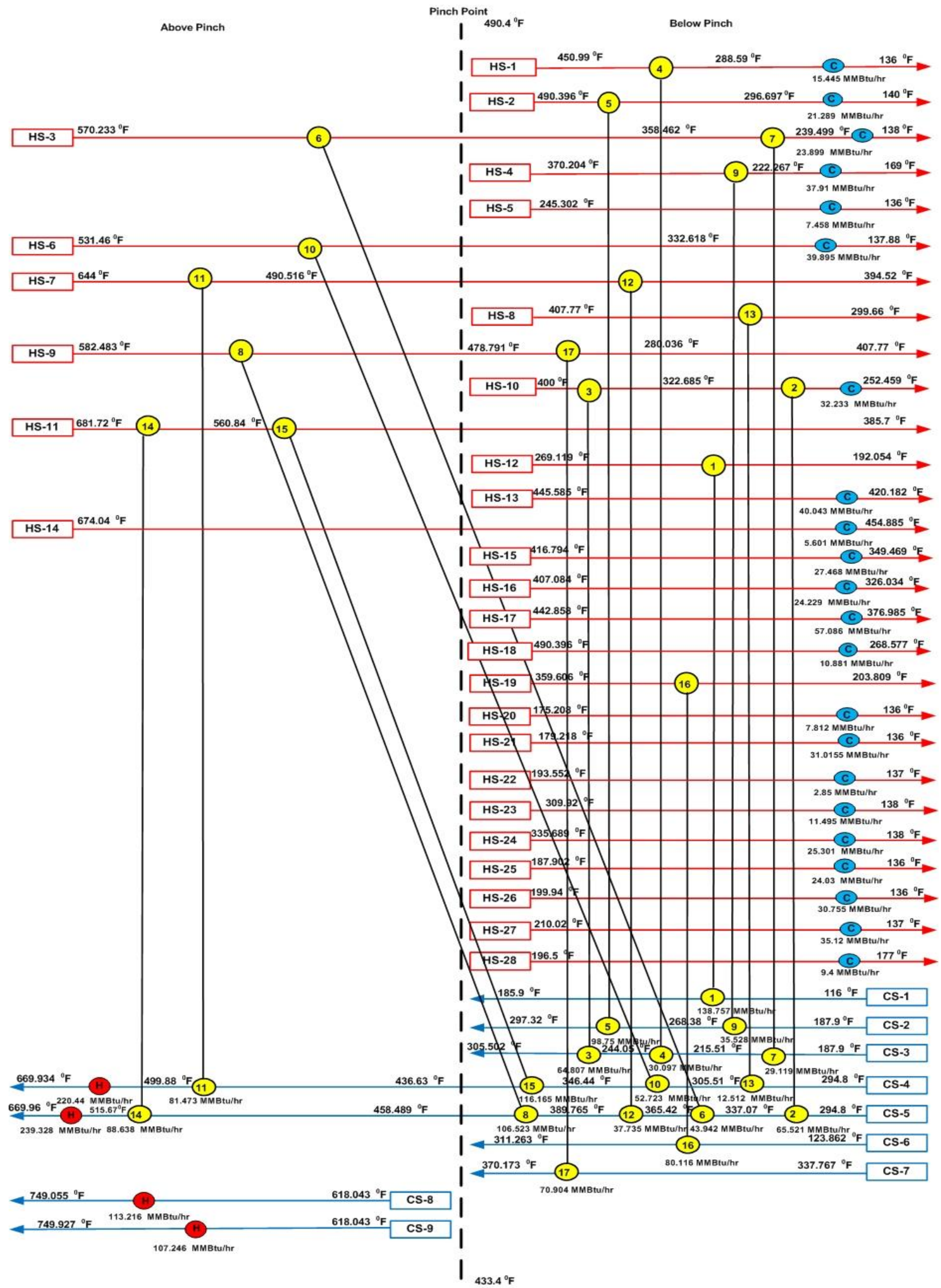
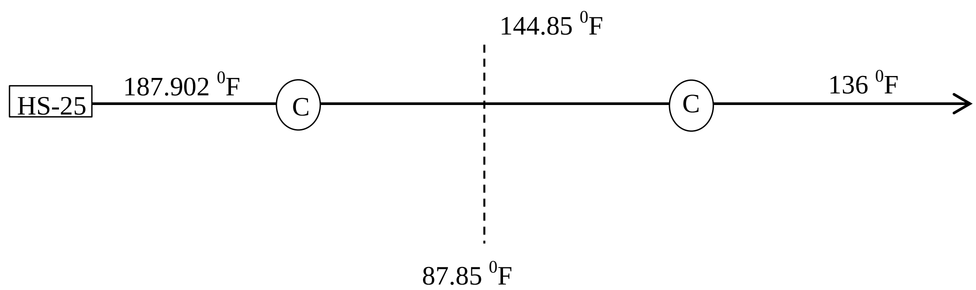
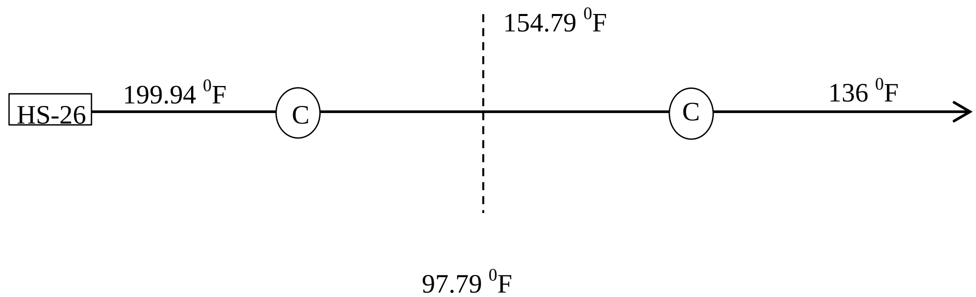


Figure 4.6: Process to Process Heat Exchanger Units that Transfer Heat Across the Pinch from Above to Below the Pinch in the Existing Design.

1. Heat Exchanger (HE-18cu)



2. Heat Exchanger (HE-19cu)



3. Heat Exchanger (HE-35cu)

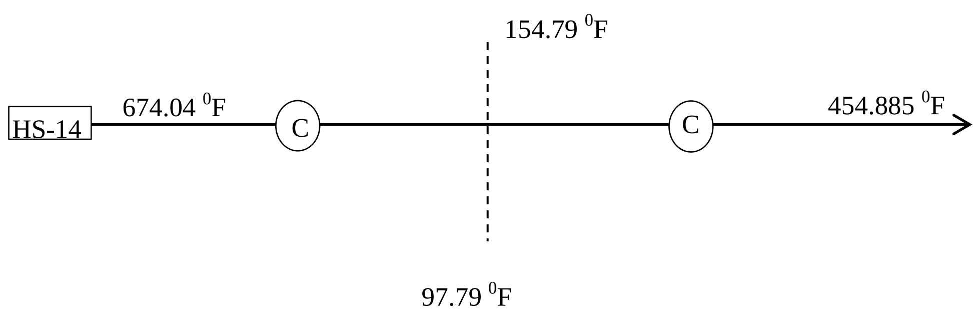


Figure 4.7: Cold Utility Units that are Used Inappropriately in the Existing Design Removing Heat from Above the Pinch.

CHAPTER 5

MATHMATICAL PROGRAM

5.1 Introduction

Mathematical program or optimization model is deal with the formulation, solution, and analysis of optimization problems or mathematical programs. The analysis and solution of an optimization problem may involve graphical, algebraic, or computer-aided tools.

Mathematical program represent selections of the problem as variables of the decision and seek values that minimize or maximize objective functions of the decision variables subject to constraints on values of the variable expressing the limits on possible choices of the decision.

A mathematical program can be represented by:

Objective function ; Min $f(x)$

Subject to:

Equality constraints ; $h(x) = 0$

Inequality constraints ; $g(x) \leq 0$

Variable bounds ; $x_{min} \leq x \leq x_{max}$

The vector x is referred to as the vector of optimization variables.

The objective function includes maximization or minimization the value of a problem but with subject to a number of constraints which are in the form of equality (material and energy balances) and inequality (e.g., the quantity of certain pollutants should be below) expressions.

5.2 LP Transshipment Model

The target of minimum utilities consumption for a heat exchanger network that has a minimum temperature approach of 57 ^{0}F can be formulated as a linear programming LP transshipment model.

A linear model is used to determine the minimum energy requirement (MER). In terms of the transshipment model hot streams are treated as source, and cold streams as destination. Heat can then be regarded as a commodity that must be transferred from the sources to the destinations through some intermediate. Figure 5.1 shows Analogy between the transshipment model and HEN.

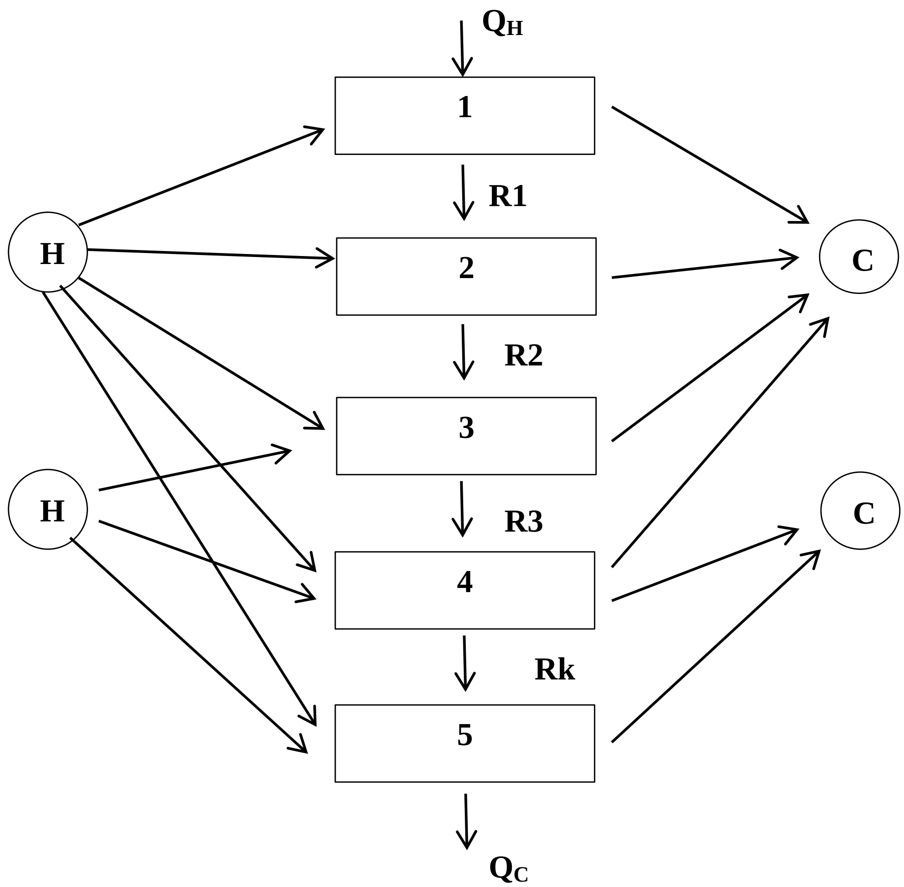


Figure 5.1: Analogy between the Transshipment Model and HEN [8]

Figure 5.2 shows heat flows in interval k that is based on the inlet temperatures of the process streams and of the intermediate utilities whose inlet temperatures fall within the range of the temperatures of the process streams. We assume that the intervals are numbered from the top to the bottom. We can then define the following index sets:

$Q_{C,k}$: heat content of cold stream demands heat from interval k

$Q_{H,k}$: heat content of hot stream supplies heat to interval k

Ri,k-1, Ri,k : heat residual inlet and outlet of interval k

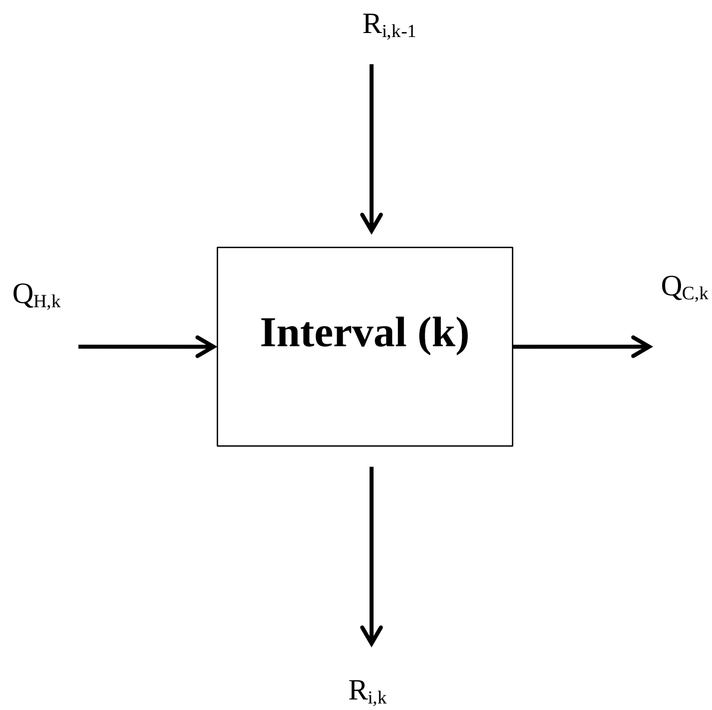


Figure 5.2: Heat Flows in Interval k

5.2.1 Formulation

The formulation includes objective function that minimizes the energy utility consumption and constraints of heat balance.

The objective of this study is to use heat integration via the transshipment linear programming model to identify the target for minimum heating and cooling utilities of the existing heat exchangers network of plant.

The formulation model is shown below:

Minimize = $Q_H^{min} + Q_C^{min}$

Subject to:

R1 - Q_H^{min} = Const.

R2 - R1 = Const.

R3 - R2 = Const.

…………

R58 - R57 = Const.

Q_C^{min} - R58 = Const.

$Q_H^{min} >= 0$, $Q_C^{min} >= 0$

R1 >= 0, … R58 >= 0

Where;

Q_H^{min} : Minimum heating utility (MMBtu/hr)

Q_C^{min} : Minimum cooling utility (MMBtu/hr)

R1 to R58: variables represent heat residual (MMBtu/hr).

5.3 Energy Target

To estimate minimum energy utilities consumption for heat exchangers network of the process, the temperature-interval diagram is constructed based on the hot and cold streams, as shown in Figure 5.3. Then, we calculated the problem table and cascade diagram which are illustrated by Table 5.1 and Figure 5.4 respectively.

Table 5.1: Problem Table

Shift Temperature °F	**Interval**	**$T_{(i-1)}-T_i$** °F	**mCp_{net}** MMBtu(IT)/hr/°F	**ΔH** MMBtu(IT)/hr	
788.427					
	1	0.872	-1.3104	-1.1427	demand
787.555					
	2	79.095	-2.4546	-194.1443	demand
708.46					
	3	0.026	-4.0051	-0.1041	demand
708.434					
	4	51.891	-5.2934	-274.6824	demand
656.543					
	5	13.323	-2.8389	-37.8224	demand
643.22					
	6	7.68	-2.1447	-16.4711	demand
635.54					
	7	30.04	-1.8947	-56.9161	demand
605.5					
	8	61.517	-1.4169	-87.1621	demand
543.983					
	9	12.25	-0.4013	-4.9165	demand
531.733					
	10	38.773	-0.177	-6.8647	demand
492.96					
	11	41.064	0.0583	2.3921	surplus
451.896					
	12	35.511	0.3753	13.3274	surplus
416.385					
	13	3.895	0.1253	0.4881	demand
412.49					
	14	3.817	0.2699	1.0301	surplus
408.673					
	15	1.588	-1.9181	-3.0459	demand
407.085					
	16	2.727	-0.3421	-0.9329	demand
404.358					
	17	22.676	0.6763	15.3358	surplus
381.682					
	18	3.388	-0.8997	-3.0482	demand
378.294					
	19	2.027	-0.4917	-0.9967	demand
376.267					
	20	6.997	1.6963	11.8688	surplus
369.27					
	21	0.686	0.7965	0.5464	demand
368.584					
	22	7.084	1.5395	10.9056	surplus
361.5					
	23	5.48	2.6413	14.4742	surplus
356.02					
	24	6.257	2.1635	13.5368	surplus
349.763					
	25	2.563	1.736	4.4493	surplus
347.2					
	26	3.198	1.0418	3.3316	surplus
344.002					
	27	5.517	-0.0114	-0.0629	demand
338.485					
	28	4.634	-1.0298	-4.7721	demand
333.851					
	29	0.551	-2.2728	-1.2523	demand
333.3					
	30	1.596	0.5661	0.9035	demand
331.704					
	31	1.627	1.2453	2.0261	surplus

330.077					
	32	8.971	1.0904	9.7821	surplus
321.106					
	33	10.137	1.6046	16.2659	surplus
310.969					
	34	13.78	1.1966	16.4894	surplus
297.189					
	35	9.655	1.3246	12.789	surplus
287.534					
	36	16.114	0.5816	9.3719	surplus
271.42					
	37	10.26	0.6484	6.6526	surplus
261.16					
	38	30.541	0.5327	16.2682	surplus
230.619					
	39	4.798	2.3332	11.1945	surplus
225.821					
	40	11.862	2.6838	31.8355	surplus
213.959					
	41	7.157	1.582	11.3225	surplus
206.802					
	42	35.282	1.6503	58.2241	surplus
171.52					
	43	6.211	2.1312	13.2366	surplus
165.309					
	44	2.947	1.617	4.7652	surplus
162.362					
	45	0.922	2.0445	1.885	surplus
161.44					
	46	3.44	2.5245	8.6841	surplus
158					
	47	2.948	3.0065	8.863	surplus
155.052					
	48	0.552	3.0568	1.6874	surplus
154.5					
	49	0.946	5.0023	4.7322	surplus
153.554					
	50	4.152	3.2018	13.294	surplus
149.402					
	51	8.684	3.6648	31.8253	surplus
140.718					
	52	2.218	4.3593	9.669	surplus
138.5					
	53	1.792	3.8773	6.9482	surplus
136.708					
	54	6.208	4.0765	25.307	surplus
130.5					
	55	29	3.3973	98.522	surplus
101.5					
	56	2	3.2352	6.4703	surplus
99.5					
	57	0.12	2.8161	0.3379	demand
99.38					
	58	0.88	2.5808	2.2711	surplus
98.5					
	59	1	2.0495	2.0495	surplus
97.5					

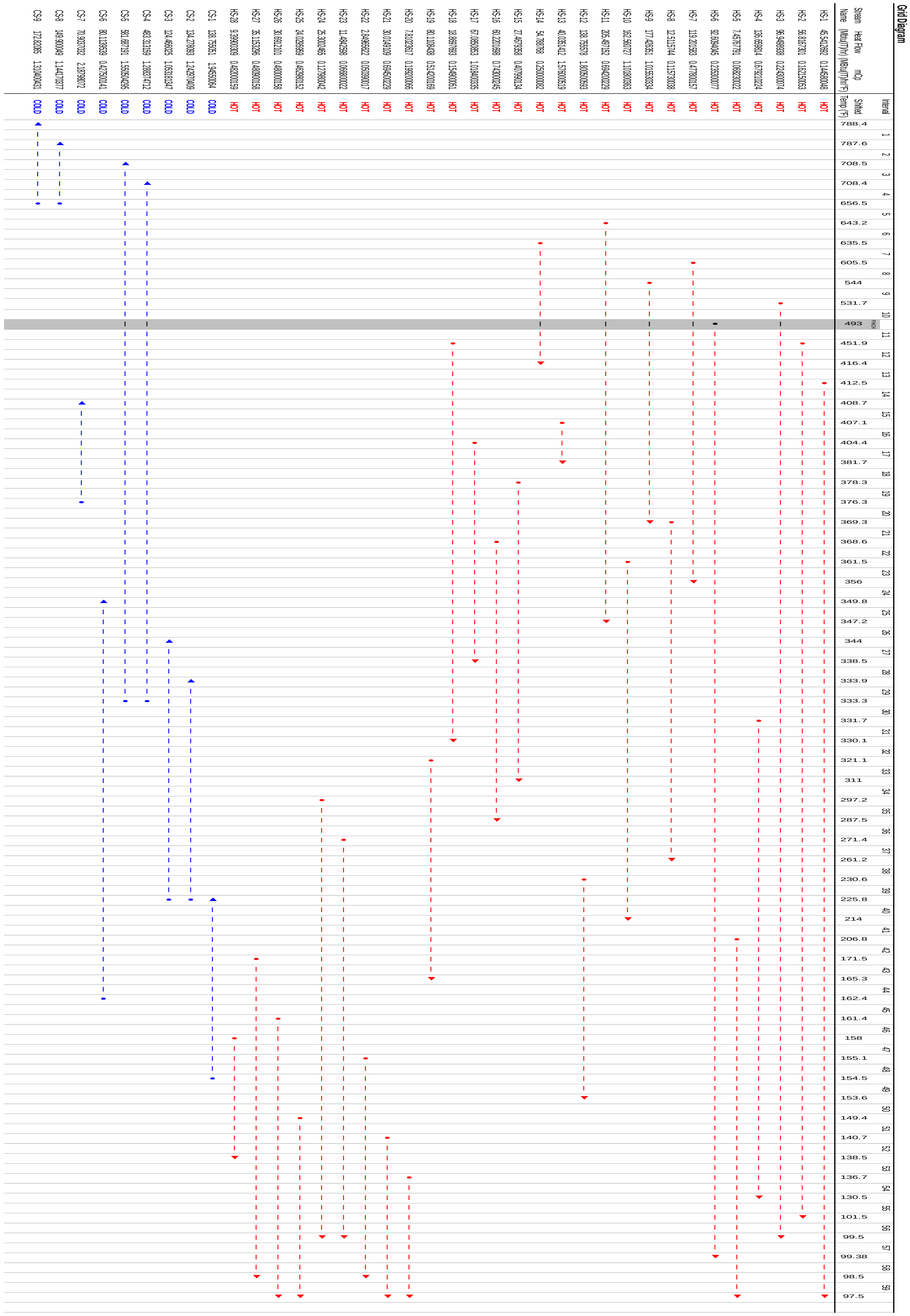


Figure 5.3: Grid Diagram of Temperature-interval

	▼	$Q_{H,min}$ =	0
k=1	-1.14267		
	▼	R1 =	-1.14267
k=2	-194.144		
	▼	R2 =	-195.29
….	-0.10413		
	▼		-195.39
	-239.338		
	▼		-434.73
	-30.7092		
	▼		-465.44
	-2.15111		
	▼		-467.59
	-56.9161		
	▼		-524.51
	-87.1621		
	▼		-611.67
	-4.91652		
	▼		-616.58
	-6.86469		
	▼		-623.45
	2.392058		
	▼		-621.06
	13.32735		
	▼		-607.73
	0.488051		
	▼		-607.24
	1.458712		
	▼		-605.78
	5.033722		
	▼		-600.75
	44.92628		
	▼		-555.82
	4.728028		
	▼		-551.09
	-3.04818		
	▼		-554.14
	-4.43718		
	▼		-558.58
	-0.95457		
	▼		-559.53
	-4.59403		
	▼		-564.13
	2.372077		
	▼		-561.76
	0.652394		
	▼		-561.1
	19.08183		
	▼		-542.02
	12.80471		
	▼		-529.22
	3.057364		
	▼		-526.16
	1.838644		
	▼		-524.32
	0.306207		
	▼		-524.02
	3.155197		
	▼		-520.86
	-1.46393		
	▼		-522.32
	0.063125		

		-522.26
-0.68008		-522.94
3.740345		-519.2
16.48942		-502.71
12.78901		-489.92
9.371889		-480.55
6.652576		-473.9
16.26825		-457.63
38.87061		-418.76
8.812911		-409.95
1.274907		-408.67
56.60524		-352.07
13.23658		-338.83
6.255982		-332.57
7.21351		-325.36
7.602747		-317.76
3.938753		-313.82
3.441345		-310.38
9.094416		-301.28
2.826563		-298.46
5.353855		-293.1
3.461823		-289.64
4.705294		-284.94
16.30609		-268.63
98.52202		-170.11
6.470322		-163.64
0.33793		-163.3
2.271087	R58 =	-161.03
2.049491	$Q_{C,min}$ =	-158.98

Figure 5.4: Cascade Diagram

5.4 Results and Discussion

The minimum heating and cooling utility were determined to be 623.44 MMBtu/hr and 464.46 MMBtu/hr, respectively. These results were consistent with those resulted from heat integration using the graphic method shown in Figure 4.5 via Hint Software and from LINGO linear programming with the set formulations shown in appendix 'B'. After heat integration, the heating and cooling utility was reduced by 8.35% and 10.89%.

CHAPTER 6

RETROFIT DESIGN USING NETWORK PINCH METHOD

6.1 Introduction

Retrofit target technique for existing heat exchangers network based on the minimum approach temperature to achieve the reduction of utility consumption. The optimum approach temperature of 57 ^{0}F for minimum energy consumption and maximizing energy recovery of the process is known via the composite curve. The scope for saving of energy and the target of area is set of the performance of the existing network. The recovery of energy of the existing network is constrained by the bottleneck of the process. The network pinch is overcome by a topology change. The topology changes is achieved in the diagnosis stage, which overcome the network pinch and increase conservation of energy via shifting heat from below to above the network pinch. The modifications include resequencing, repiping, addition of new heat exchangers and spilt of stream.

6.2 Diagnosis Stage

In this stage, the aim is to classify design options that overcomes the network pinch and increases the heat recovery of network. The search of topology change is sequential and requires user interaction to evaluate the modifications. The best few design options are identified at this stage. These are later taken to the optimizer to find the optimum network topology. To carry out the network pinch method, a minimum approach temperature needs to be specified. The area efficiency method was used for targeting therefore the optimum value of the minimum approach temperature obtained from it, which is 57 ^{0}F, will be used for the retrofit design in this study. The design objective in the diagnosis stage is set for minimum energy consumption. The search of modifications includes resequencing, repiping, addition of new heat exchangers and spilt of stream. The modifications are investigated in that order since that is the order that requires less changes and complications in the design. The procedure is repeated at each modification step and due to different options generated during each modification, the best few design options are developed, which are then taken to the optimizer and compared. Figure 6.1 summaries the steps taken in diagnosis stage for the different design options.

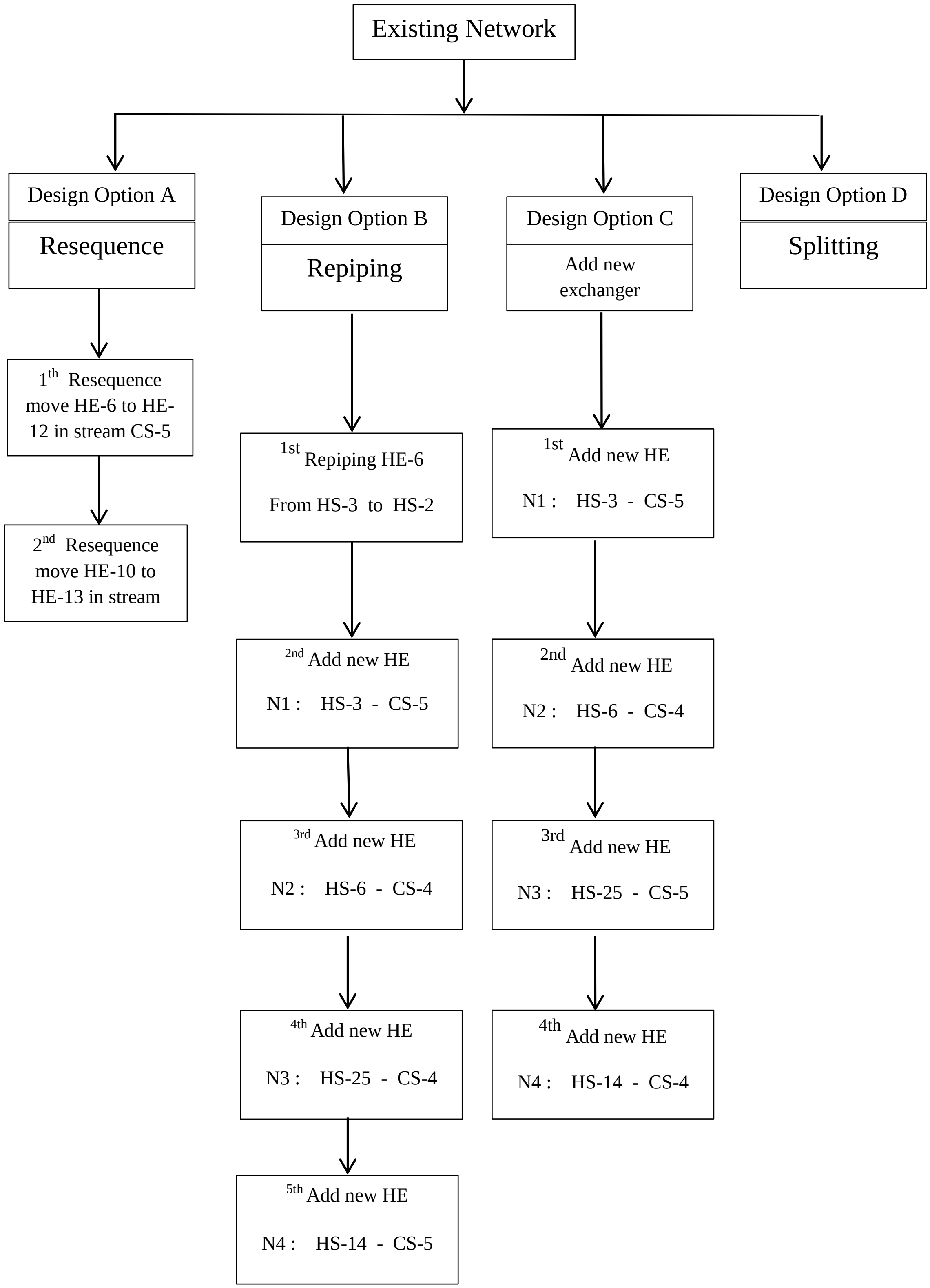


Figure 6.1: The Design Solution Search Tree for Diagnosis Stage of the Plant HEN

6.2.1 Design Option (A)

This option starts with resequencing of the existing heat exchanger of the heat exchangers network to obtain maximum energy savings.

1. The search for new modification of option (A) via resequencing of heat exchanger (HE-6) moving it after heat exchanger (HE-12) in stream CS-5. Optimization results did not give any energy saving over the existing network.

2. Further attempt was made to resequence heat exchanger (HE-10) by moving it after heat exchanger (HE-13) in stream CS-4. The modification did not give any better result over existing network.

The design option (A) amounts to no beneficial improvement over existing design, hence, the search is stopped.

6.2.2 Design Option (B)

This design option starts with repiping of the existing heat exchanger network. The modifications result in energy saving via relocation of heat exchanger, removal of cold utility exchanger and addition of process-to-process heat exchangers. Figure 6.2 summarizes the results of design option (B).

1. Repiping of heat exchanger (HE-6) from hot stream HS-3 to hot stream HS-2. This gives energy saving of 4.08 % in the cold utility over the existing network.

2. The trade-off between energy and capital cost comes to play in this option. Addition of new heat exchanger (N1) between streams HS-3 and CS-5 gives 3.13% savings in the hot utility over the existing design.

3. Addition of new heat exchanger (N2) between streams HS-6 and CS-4 gives energy savings 1.47% hot utility and 1.93% cold utility.

4. Add new heat exchanger (N3) between streams HS-25 and CS-4 saves 3.29% hot utility and 4.22% cold utility over the existing heat exchanger network.

5. Further addition of new heat exchanger (N4) between streams HS-14 and CS-5 results in energy saving of 0.746% hot utility and 0.999% cold utility over the existing design.

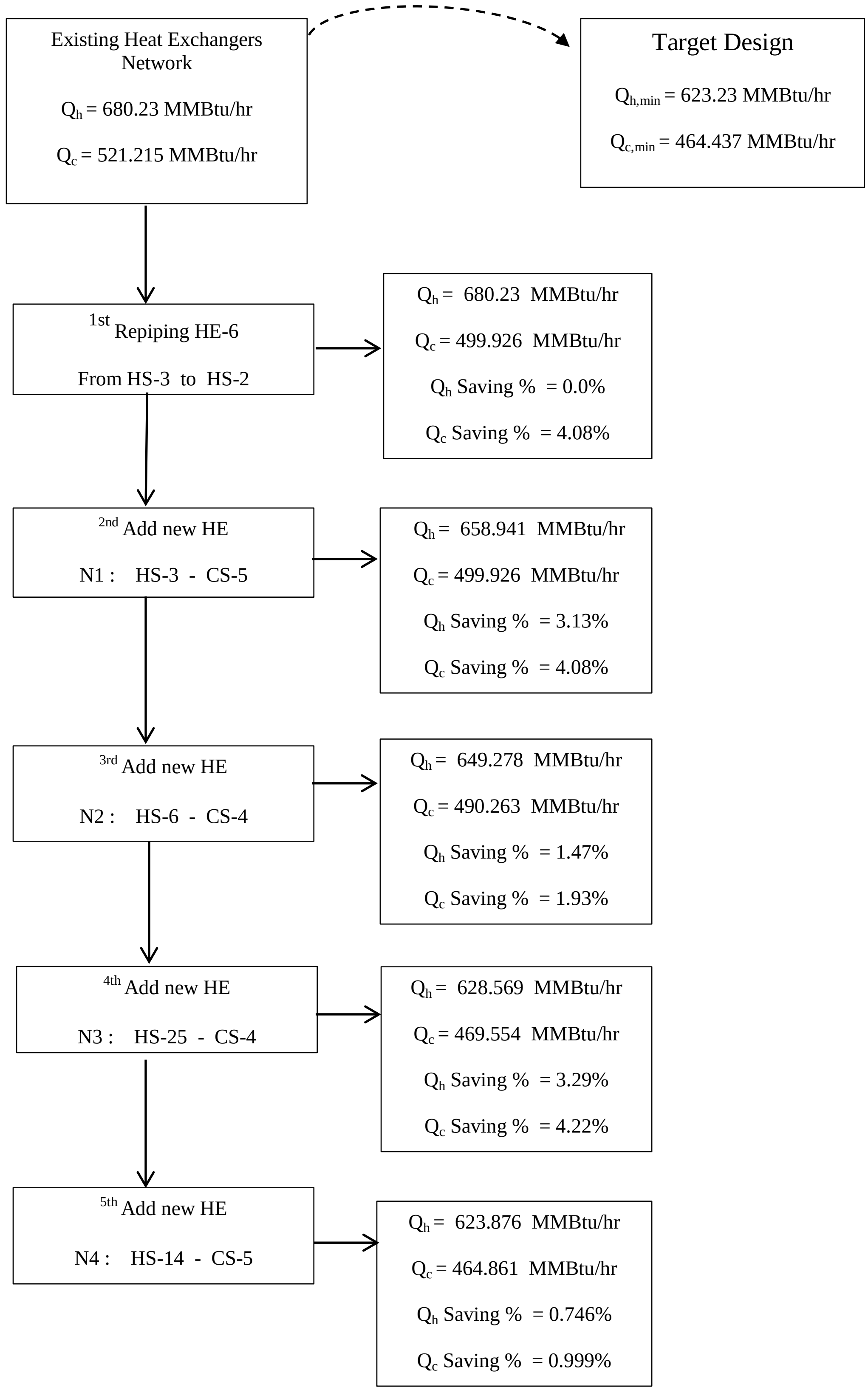


Figure 6.2: Modification of Existing Network for Design Option B

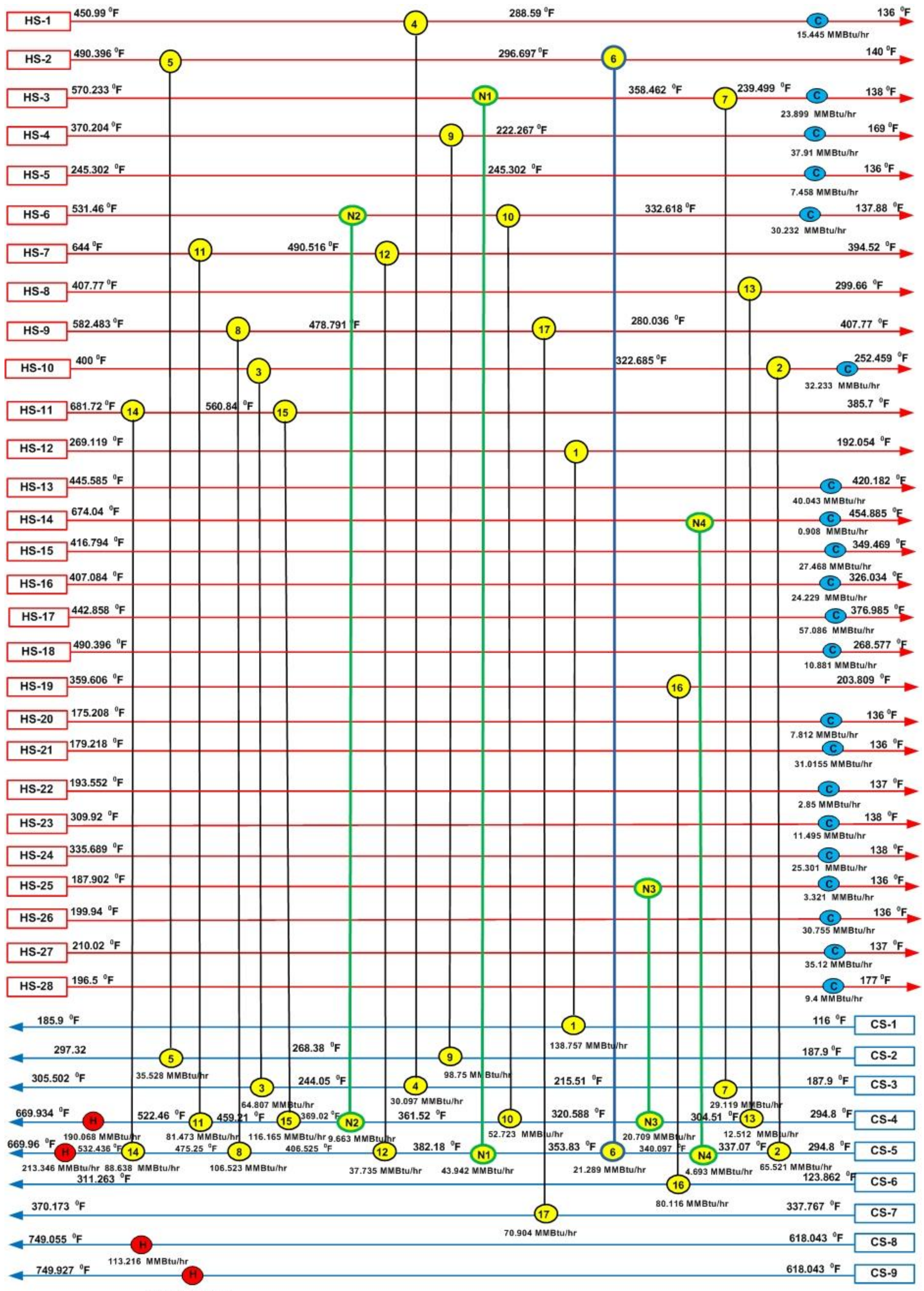


Figure 6.3: Grid Diagram of Design Option B

6.2.3 Design Option (C)

This option starts with addition of new heat exchangers to the existing heat exchangers network. This modification gave maximum energy saving via adding four heat exchangers and remove cold utility exchanger. Figure 6.4 shows the steps of the design option (C).

1. Add new heat exchanger (N1) between streams HS-3 and CS-5 so this gives energy conservation of the existing network about 2.63% hot utility and 3.43% cold utility.
2. Add new heat exchanger (N2) between streams HS-6 and CS-4 so this gives energy conservation of the existing network about 1.46% hot utility and 1.92% cold utility.
3. Add new heat exchanger (N3) between streams HS-25 and CS-5 so this gives energy conservation of the existing network about 3.17% hot utility and 4.19% cold utility.
4. Add new heat exchanger (N4) between streams HS-14 and CS-4 so this gives energy conservation of the existing network about 0.88% hot utility and 1.18% cold utility.

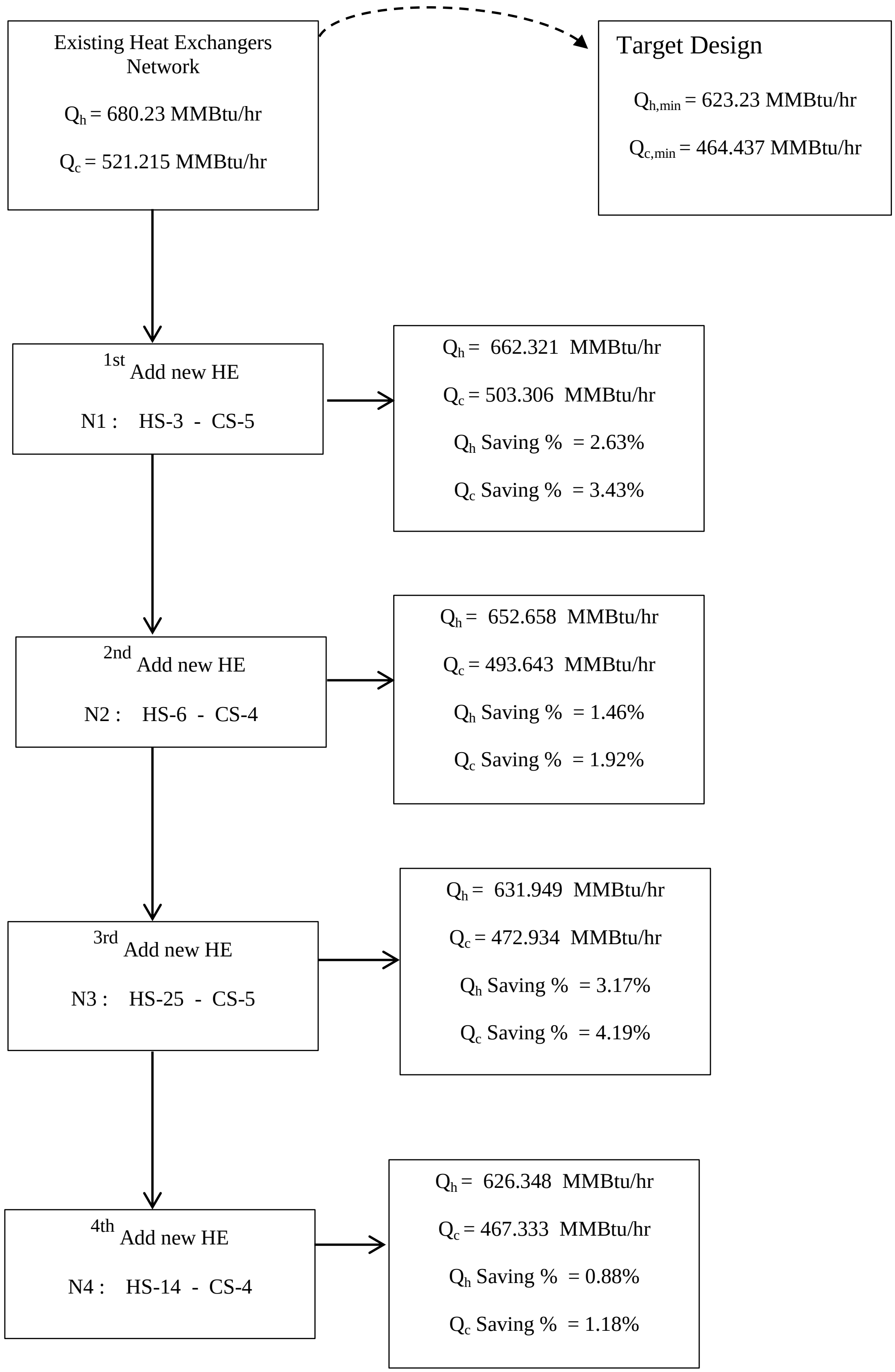


Figure 6.4: Modification of Existing Network for Design Option C

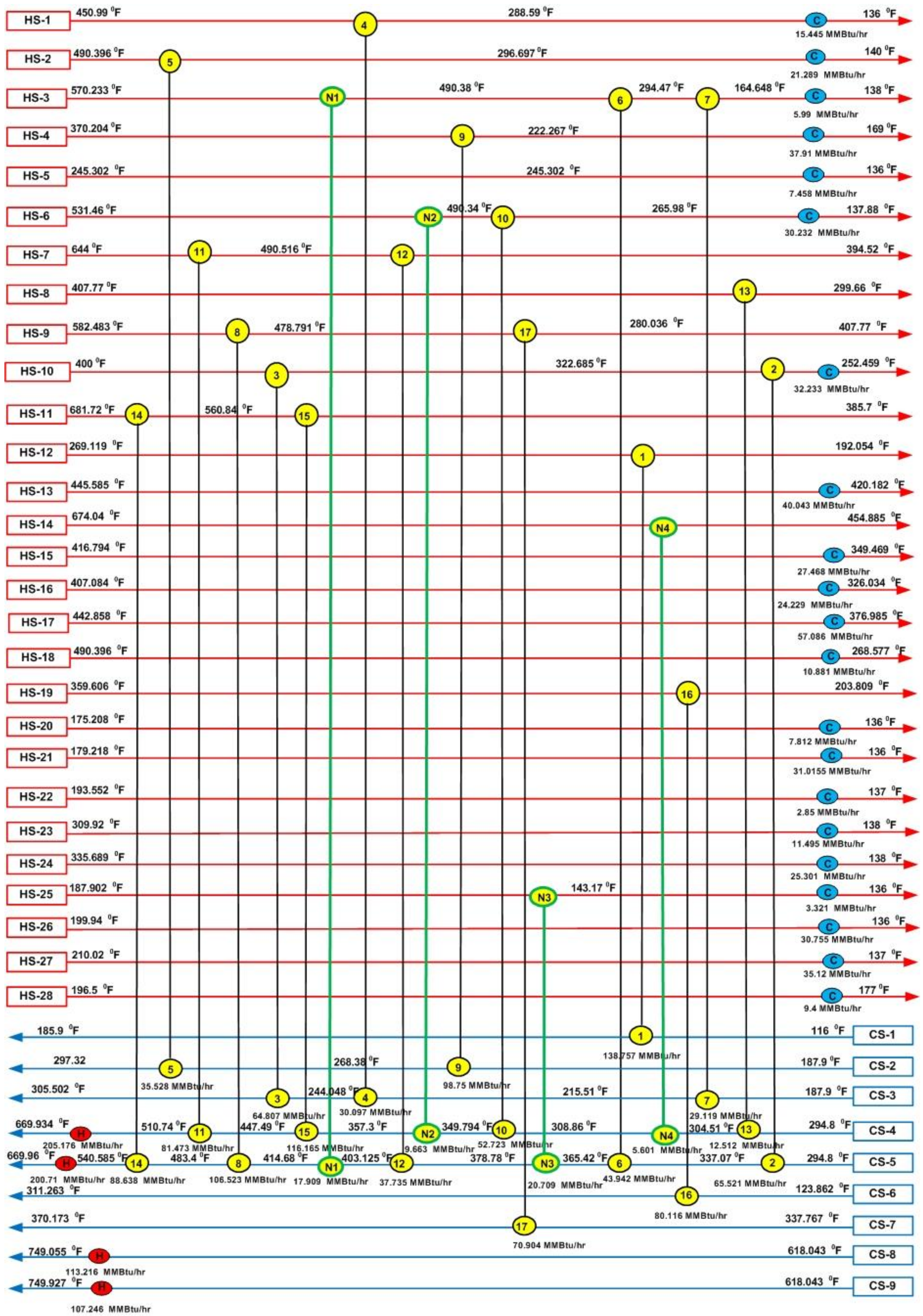


Figure 6.5: Grid Diagram of Design Option C

6.2.4 Design Option (D)

This option starts by splitting of stream of the heat exchangers network to obtain maximum energy savings. This design option results in evolution of design that is difficult to control and expensive due to extra cost of modification of pumps, piping and valves. Therefore, the search is stopped at this option (D).

6.3 Comparison and Evaluation Design Options

The results of trade-off between utility demands and number of units in the retrofit and existing designs are presented in Table 6.1.

Table 6.1: Comparison of Design Options with Existing Design

Design	Hot utility	Cold utility	No. of units
Existing	680.23 MMBtu/hr	521.215 MMBtu/hr	43
Option A	-	-	-
Option B	623.876 MMBtu/hr	464.861 MMBtu/hr	47
Option C	626.348 MMBtu/hr	467.333 MMBtu/hr	47
Option D	-	-	-

It is observed from the Table 6.1 that both the hot and cold utility demands in design options B and C are reduced when compared with existing design. However, the price paid for this reduction in utility demand is invested in addition heat exchanger units.

Since design options B and C require the same number of heat exchanger unit, a close examination of the two options revealed that option B could be a preferred candidate over option C. Nevertheless, the choice of better design between option B and existing design is difficult at this stage. Economic evaluation of the two designs will be required before a preferred option could be selected.

CHAPTER 7

CONCLUSIONS AND RECOMMENDATIONS

7.1 Discussions and Conclusions

The heat integration of the existing heat exchangers network of crude distillation units is studied in this thesis. The existing heat exchangers network consists of twenty-eight hot and nine cold streams. There are forty-three units used in the existing design seventeen of which are process-to-process heat exchangers, twenty-two cold utility units that include four water coolers, twelve air coolers and six boiler feed water coolers, and four hot utility units that used four furnaces. The plant uses light crude as feedstock.

The pinch technology is based on thermodynamic principles. Knowing the thermodynamic data, the scope of heat recovery and energy target can be set using the composite curve. The composite curve defines the cumulative heat source and sink of the process. The economic data include energy and capital cost and requires carrying out the pinch analysis of process.

Economic evaluation of existing design and the retrofit is shown in Table 7.1. A reduction in hot utility demand from 680.23 MMBtu/hr in the existing process to 623.455 MMBtu/hr in the new network is observed. This translates to 8.35% hot utility saving. The cold utility demand also reduced from 521.215 MMBtu/hr in the existing network to

464.437 MMBtu/hr in the new design. This reduction represents 10.89% of the cold utility demand.

Table 7.1: Economic Evaluation of Existing and Retrofit Design

Type	ΔTmin ^{0}F	Hot utility (MMBtu/hr)	Cold utility (MMBtu/hr)	Energy Cost ($/yr)	Capital Cost ($/yr)	Total Cost ($/yr)
The existing network	77	680.23	521.215	3.10382*E6	0.1725795*E7	0.4829625*E7
The new design network	57	623.455	464.437	2.84396*E6	0.1982385*E7	0.4826345*E7
Scope saving		56.775	56.778	0.25986*E6	- 0.025659*E7	0.3280*E4
Saving %		8.35%	10.89%	8.37%		
Payback	One year					

The energy saving in monetary terms is $256,860/yr. The capital investment of additional heat exchangers is justified by the payback period of one year to recoup the money spent. The new design is presented in Figure 7.1. It has twenty-one process-to-process heat exchangers, twenty-two cold utility units that include: four water coolers, twelve air coolers, and six boilers feed water coolers. It also has four hot utility units (four furnaces).

The retrofit design is carried out using the network pinch method. The network pinch is overcome by topology changes to shift heat from below to above the network pinch. For the existing heat exchangers network retrofit to get the optimum new design and to satisfy of the minimum energy consumption demand should be increased heat exchange area for maximum heat recovery. The design that is selected for the retrofit is design option B that has maximum heat recovery, i.e., Better energy savings about 56.354 MMBtu/hr of hot utility and 56.354 MMBtu/hr of cold utility. This option has repiping one heat exchanger (HE-6) with removing air-cooling (27cu) and adding four heat exchangers (N1, N2, N3 and N4), where is added to the network to satisfy of the minimum energy consumption demand requirements. Figure 7.1 shows the optimum design for heat exchangers network at ΔTmin= 57 ^{0}F.

In addition, evaluation of general energy conservation of plant is given in table 7.2. It is noted that energy utilities of plant before heat integration about 3515 MMBtu/hr but after heat integration 3401.222 MMBtu/hr, i.e. the scope of energy saving of all the plant about 113.778 MMBtu/hr.

Table 7.2: Evaluation of General Energy Conservation

Types	Before Heat Integration	After Heat Integration	Energy Conservation
Energy Utilities of Plant	3515 MMBtu/hr	3401.222 MMBtu/hr	113.778 MMBtu/hr
Consumption Energy of HEN	1201.445 MMBtu/hr	1087.667 MMBtu/hr	113.778 MMBtu/hr
Furnaces	680.23 MMBtu/hr	623.455 MMBtu/hr	56.775 MMBtu/hr
Boiler Feed Water	165.308 MMBtu/hr	160.615 MMBtu/hr	4.693 MMBtu/hr
Air cooling	256.6025 MMBtu/hr	225.6515 MMBtu/hr	30.951 MMBtu/hr
Cooling Water	99.305 MMBtu/hr	78.596 MMBtu/hr	20.709 MMBtu/hr
Area of Heat Exchangers Network	178876 ft^2	214412 ft^2	——
Network Diagram	• Seventeen heat exchangers, process-to-process. • Twenty-two cold utility units that include : Four water coolers Twelve air coolers Six boilers feed water coolers. • Four hot utility units that include: Four furnaces	• Twenty-one heat exchangers, process-to-process. • Twenty-two cold utility units that include : Four water coolers Twelve air coolers Six boilers feed water coolers. • Four hot utility units that include: Four furnaces.	

Maximum energy consumption of heat exchanger network of an existing plant about 1201.445 MMBtu/hr while at new design of heat exchanger network about 1087.667 MMBtu/hr i.e. the scope of energy savings about 56.775 MMBtu\hr for hot utility and for cold utility about 4.693 MMBtu/hr as boiler feed water 30.951 MMBtu/hr as air cooling and 20.709 MMBtu/hr as cooling water.

In this study, heat integration and retrofit of heat exchanger network of existing operational oil refining crude distillation unit was carried out. Pinch analyses of the existing design showed that process-to-process heat exchangers HE-6, HE-8, HE-10, HE-11 and HE-15 were transferring heat across the pinch. This violation of pinch rule offered the scope for retrofit project in the design. The network pinch method was adopted in the search for the heat exchanger network that optimized energy recovery in the process. A number of design options and modifications were subsequently carried out, of which design option B offered the most promising design solution. Economic evaluation of selected design showed hot utility and cold utility saving of 8.35% and 10.89%, respectively over the existing design. The energy cost saving of $259,860/yr was also achieved. The modifications of existing design suggested repiping of heat exchanger HE-6 and addition of four new exchangers. The cost incurred as a result of modification of existing design was $256,590 with a payback period of only one year.

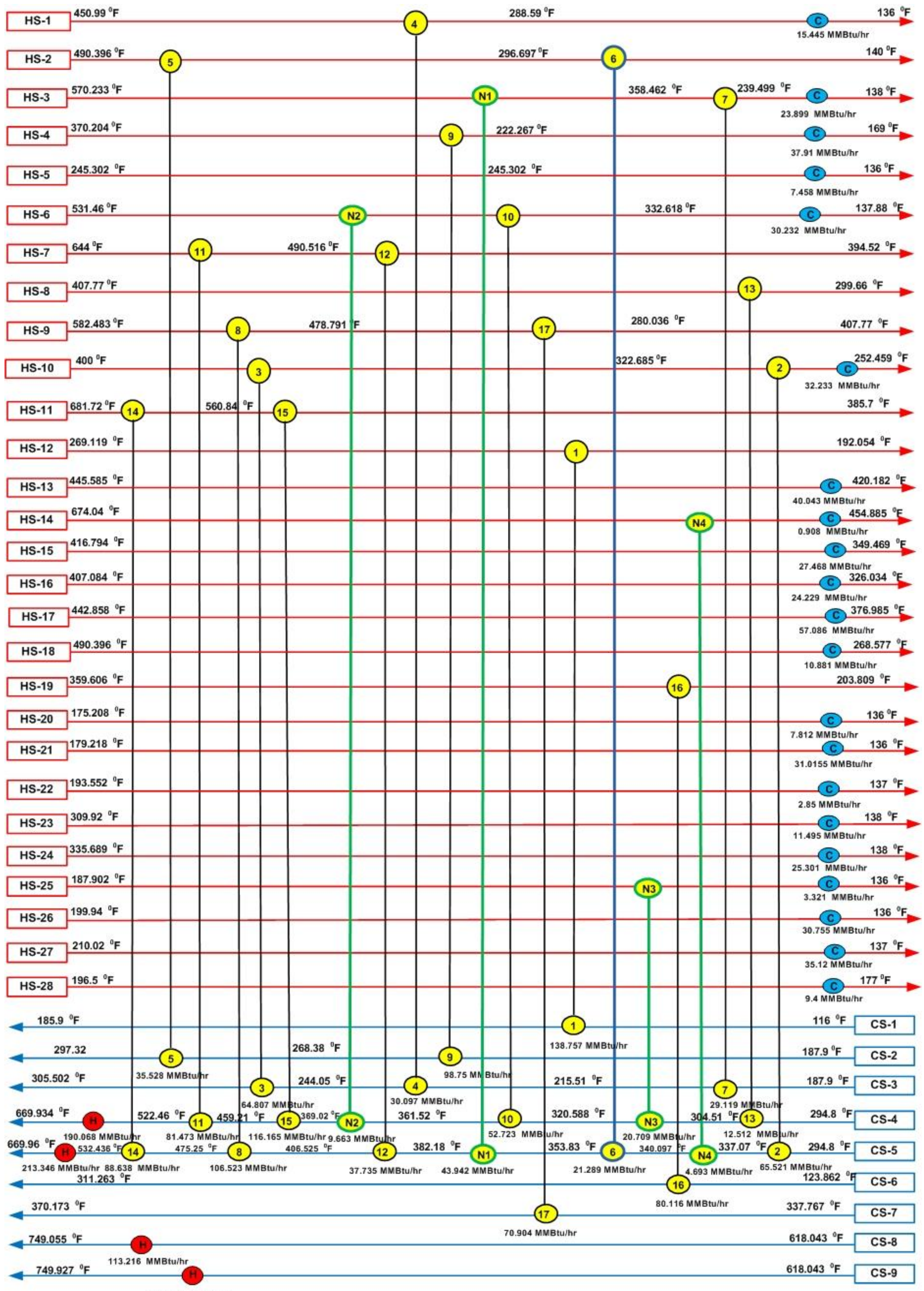


Figure 7.1: The Optimum Design for Heat Exchangers Network at ΔT_{min}= 57 ^{0}F

7.2 Recommendations and Suggestions for Further Work

- Heat transfer enhancement technique analysis can be used to reduce the additional area required.
- The study about amount of emissions caused by burning fuel (CO_2, SO_2, NOx etc.) in the furnaces for environmental protection.
- The work for integrating main fractionation and vacuum columns by using pinch analysis, will give more savings of energy.

APPENDIX A

Pinch Analysis Data

Table 4.8: Values of Range Target Plot of Total (A1-1) Cost Versus in the Region of Optimum Value

Dtmin [F]	Minimum Hot Util. [MMBtu/hr]	Minimum Cold Util. [MMBtu/hr	CC Area Target [ft^2]	Units (-)	Utility Cost [$/yr]	Capital Cost [$/yr]	Total Cost [$/yr]
1.0	464.5	305.5	0.1545E+07	45	0.2118E+07	0.5776E+07	0.7895E+07
2.0	467.3	308.3	0.1111E+07	45	0.2131E+07	0.4756E+07	0.6887E+07
3.0	470.2	311.1	0.9048E+06	45	0.2144E+07	0.4218E+07	0.6362E+07
4.0	473.0	314.0	0.7797E+06	45	0.2157E+07	0.3867E+07	0.6024E+07
5.0	475.8	316.8	0.6939E+06	45	0.2170E+07	0.3614E+07	0.5784E+07
6.0	478.7	319.7	0.6306E+06	45	0.2183E+07	0.3420E+07	0.5603E+07
7.0	481.5	322.5	0.5816E+06	45	0.2196E+07	0.3263E+07	0.5459E+07
8.0	484.4	325.3	0.5422E+06	45	0.2209E+07	0.3134E+07	0.5343E+07
9.0	487.2	328.2	0.5096E+06	45	0.2222E+07	0.3025E+07	0.5247E+07
10.0	490.0	331.0	0.4821E+06	45	0.2235E+07	0.2930E+07	0.5165E+07
11.0	492.9	333.8	0.4585E+06	45	0.2248E+07	0.2847E+07	0.5095E+07
12.0	495.7	336.7	0.4380E+06	45	0.2261E+07	0.2774E+07	0.5034E+07
13.0	498.5	339.5	0.4199E+06	45	0.2274E+07	0.2708E+07	0.4981E+07
14.0	501.4	342.4	0.4038E+06	45	0.2287E+07	0.2648E+07	0.4935E+07
15.0	504.2	345.2	0.3893E+06	45	0.2300E+07	0.2594E+07	0.4893E+07
16.0	507.1	348.0	0.3763E+06	45	0.2313E+07	0.2544E+07	0.4857E+07
17.0	509.9	350.9	0.3644E+06	45	0.2326E+07	0.2498E+07	0.4824E+07
18.0	512.7	353.7	0.3535E+06	45	0.2338E+07	0.2456E+07	0.4794E+07
19.0	515.6	356.6	0.3435E+06	45	0.2351E+07	0.2416E+07	0.4768E+07
20.0	518.4	359.4	0.3343E+06	45	0.2364E+07	0.2379E+07	0.4744E+07
21.0	521.3	362.2	0.3257E+06	45	0.2377E+07	0.2345E+07	0.4722E+07
22.0	524.1	365.1	0.3178E+06	45	0.2390E+07	0.2312E+07	0.4703E+07
23.0	526.9	367.9	0.3103E+06	45	0.2403E+07	0.2282E+07	0.4685E+07
24.0	529.8	370.8	0.3033E+06	45	0.2416E+07	0.2253E+07	0.4669E+07
25.0	532.6	373.6	0.2968E+06	45	0.2429E+07	0.2225E+07	0.4654E+07
26.0	535.4	376.4	0.2906E+06	45	0.2442E+07	0.2199E+07	0.4641E+07
27.0	538.3	379.3	0.2848E+06	45	0.2455E+07	0.2175E+07	0.4630E+07

28.0	541.1	382.1	0.2793E+06	45	0.2468E+07	0.2151E+07	0.4619E+07	
29.0	544.0	384.9	0.2741E+06	45	0.2481E+07	0.2128E+07	0.4609E+07	
30.0	546.8	387.8	0.2691E+06	45	0.2494E+07	0.2107E+07	0.4601E+07	
31.0	549.6	390.6	0.2644E+06	45	0.2507E+07	0.2086E+07	0.4593E+07	
32.0	552.5	393.5	0.2599E+06	45	0.2520E+07	0.2066E+07	0.4586E+07	
33.0	555.3	396.3	0.2557E+06	45	0.2533E+07	0.2047E+07	0.4580E+07	
34.0	558.2	399.1	0.2516E+06	45	0.2546E+07	0.2029E+07	0.4575E+07	
35.0	561.0	402.0	0.2477E+06	45	0.2559E+07	0.2012E+07	0.4570E+07	
36.0	563.8	404.8	0.2440E+06	45	0.2572E+07	0.1995E+07	0.4566E+07	
37.0	566.7	407.7	0.2404E+06	45	0.2584E+07	0.1978E+07	0.4563E+07	
* 38.0	569.5	410.5	0.2369E+06	45	0.2597E+07	0.1963E+07	0.4560E+07	*
39.0	572.4	413.3	0.2336E+06	46	0.2610E+07	0.1966E+07	0.4577E+07	
40.0	575.2	416.2	0.2303E+06	46	0.2623E+07	0.1951E+07	0.4575E+07	
41.0	578.0	419.0	0.2271E+06	46	0.2636E+07	0.1936E+07	0.4573E+07	
42.0	580.9	421.9	0.2241E+06	46	0.2649E+07	0.1922E+07	0.4571E+07	
43.0	583.7	424.7	0.2211E+06	46	0.2662E+07	0.1908E+07	0.4570E+07	
44.0	586.5	427.5	0.2183E+06	46	0.2675E+07	0.1894E+07	0.4569E+07	
45.0	589.4	430.4	0.2155E+06	46	0.2688E+07	0.1881E+07	0.4569E+07	
46.0	592.2	433.2	0.2129E+06	46	0.2701E+07	0.1868E+07	0.4569E+07	
47.0	595.1	436.0	0.2103E+06	46	0.2714E+07	0.1856E+07	0.4570E+07	
48.0	597.9	438.9	0.2078E+06	46	0.2727E+07	0.1844E+07	0.4571E+07	
49.0	600.7	441.7	0.2053E+06	47	0.2740E+07	0.1849E+07	0.4589E+07	
50.0	603.6	444.6	0.2028E+06	47	0.2753E+07	0.1837E+07	0.4590E+07	
51.0	606.4	447.4	0.2004E+06	47	0.2766E+07	0.1825E+07	0.4591E+07	
52.0	609.3	450.2	0.1980E+06	47	0.2779E+07	0.1813E+07	0.4592E+07	
53.0	612.1	453.1	0.1957E+06	47	0.2792E+07	0.1802E+07	0.4594E+07	
54.0	614.9	455.9	0.1935E+06	47	0.2805E+07	0.1790E+07	0.4595E+07	
55.0	617.8	458.8	0.1913E+06	47	0.2818E+07	0.1779E+07	0.4597E+07	
56.0	620.6	461.6	0.1892E+06	47	0.2831E+07	0.1769E+07	0.4600E+07	
57.0	623.5	464.4	0.1872E+06	47	0.2844E+07	0.1758E+07	0.4602E+07	
58.0	626.3	467.3	0.1852E+06	47	0.2857E+07	0.1748E+07	0.4605E+07	
59.0	629.1	470.1	0.1833E+06	47	0.2870E+07	0.1738E+07	0.4608E+07	
60.0	632.0	473.0	0.1814E+06	47	0.2883E+07	0.1728E+07	0.4611E+07	
61.0	634.8	475.8	0.1795E+06	47	0.2896E+07	0.1719E+07	0.4615E+07	
62.0	637.6	478.6	0.1778E+06	47	0.2909E+07	0.1710E+07	0.4618E+07	
63.0	640.5	481.5	0.1760E+06	47	0.2922E+07	0.1700E+07	0.4622E+07	
64.0	643.3	484.3	0.1743E+06	47	0.2935E+07	0.1692E+07	0.4626E+07	
65.0	646.2	487.1	0.1727E+06	47	0.2948E+07	0.1683E+07	0.4631E+07	
66.0	649.0	490.0	0.1711E+06	47	0.2961E+07	0.1674E+07	0.4635E+07	
67.0	651.8	492.8	0.1695E+06	47	0.2974E+07	0.1666E+07	0.4640E+07	
68.0	654.7	495.7	0.1680E+06	47	0.2987E+07	0.1658E+07	0.4645E+07	
69.0	657.5	498.5	0.1665E+06	47	0.3000E+07	0.1650E+07	0.4650E+07	
70.0	660.4	501.3	0.1650E+06	47	0.3013E+07	0.1642E+07	0.4655E+07	

71.0	663.2	504.2	0.1636E+06	47	0.3026E+07	0.1634E+07	0.4660E+07
72.0	666.0	507.0	0.1622E+06	47	0.3039E+07	0.1627E+07	0.4666E+07
73.0	668.9	509.9	0.1608E+06	47	0.3052E+07	0.1619E+07	0.4671E+07
74.0	671.7	512.7	0.1595E+06	47	0.3065E+07	0.1612E+07	0.4677E+07
75.0	674.6	515.5	0.1582E+06	47	0.3078E+07	0.1605E+07	0.4683E+07
76.0	677.4	518.4	0.1570E+06	47	0.3091E+07	0.1598E+07	0.4689E+07
77.0	680.2	521.2	0.1557E+06	47	0.3104E+07	0.1591E+07	0.4695E+07

Table 4.9: Values of Range Target Plot of Total (A1-2) Cost Versus in the Region of Optimum Value

Dtmin [F]	Minimum Hot Util. [MMBtu/hr]	Minimum Cold Util. [MMBtu/hr	1-2ST Area Target [ft^2]	N Shells (-)	Utility Cost [$/yr]	Capital Cost [$/yr]	Total Cost [$/yr]
1.0	464.5	305.5	0.2073E+07	507	0.2118E+07	0.1740E+08	0.1952E+08
2.0	467.3	308.3	0.1486E+07	363	0.2131E+07	0.1252E+08	0.1465E+08
3.0	470.2	311.1	0.1207E+07	296	0.2144E+07	0.1022E+08	0.1236E+08
4.0	473.0	314.0	0.1038E+07	254	0.2157E+07	0.8804E+07	0.1096E+08
5.0	475.8	316.8	0.9223E+06	226	0.2170E+07	0.7846E+07	0.1002E+08
6.0	478.7	319.7	0.8369E+06	206	0.2183E+07	0.7149E+07	0.9332E+07
7.0	481.5	322.5	0.7707E+06	190	0.2196E+07	0.6601E+07	0.8798E+07
8.0	484.4	325.3	0.7176E+06	177	0.2209E+07	0.6160E+07	0.8369E+07
9.0	487.2	328.2	0.6736E+06	166	0.2222E+07	0.5792E+07	0.8014E+07
10.0	490.0	331.0	0.6366E+06	157	0.2235E+07	0.5485E+07	0.7720E+07
11.0	492.9	333.8	0.6048E+06	149	0.2248E+07	0.5218E+07	0.7466E+07
12.0	495.7	336.7	0.5771E+06	142	0.2261E+07	0.4985E+07	0.7246E+07
13.0	498.5	339.5	0.5528E+06	136	0.2274E+07	0.4782E+07	0.7056E+07
14.0	501.4	342.4	0.5311E+06	131	0.2287E+07	0.4606E+07	0.6892E+07
15.0	504.2	345.2	0.5116E+06	126	0.2300E+07	0.4441E+07	0.6741E+07
16.0	507.1	348.0	0.4941E+06	122	0.2313E+07	0.4299E+07	0.6611E+07
17.0	509.9	350.9	0.4781E+06	118	0.2326E+07	0.4165E+07	0.6490E+07
18.0	512.7	353.7	0.4634E+06	114	0.2338E+07	0.4038E+07	0.6376E+07
19.0	515.6	356.6	0.4499E+06	111	0.2351E+07	0.3930E+07	0.6281E+07
20.0	518.4	359.4	0.4375E+06	108	0.2364E+07	0.3827E+07	0.6191E+07
21.0	521.3	362.2	0.4259E+06	105	0.2377E+07	0.3728E+07	0.6106E+07
22.0	524.1	365.1	0.4152E+06	102	0.2390E+07	0.3634E+07	0.6025E+07
23.0	526.9	367.9	0.4051E+06	100	0.2403E+07	0.3557E+07	0.5960E+07
24.0	529.8	370.8	0.3957E+06	97	0.2416E+07	0.3469E+07	0.5886E+07
25.0	532.6	373.6	0.3869E+06	95	0.2429E+07	0.3398E+07	0.5827E+07

26.0	535.4	376.4	0.3786E+06	93	0.2442E+07	0.3330E+07	0.5772E+07
27.0	538.3	379.3	0.3707E+06	91	0.2455E+07	0.3263E+07	0.5718E+07
28.0	541.1	382.1	0.3633E+06	89	0.2468E+07	0.3199E+07	0.5667E+07
29.0	544.0	384.9	0.3563E+06	87	0.2481E+07	0.3137E+07	0.5618E+07
30.0	546.8	387.8	0.3496E+06	85	0.2494E+07	0.3077E+07	0.5571E+07
31.0	549.6	390.6	0.3432E+06	84	0.2507E+07	0.3031E+07	0.5538E+07
32.0	552.5	393.5	0.3372E+06	82	0.2520E+07	0.2974E+07	0.5494E+07
33.0	555.3	396.3	0.3314E+06	81	0.2533E+07	0.2931E+07	0.5464E+07
34.0	558.2	399.1	0.3259E+06	79	0.2546E+07	0.2877E+07	0.5423E+07
35.0	561.0	402.0	0.3206E+06	78	0.2559E+07	0.2837E+07	0.5396E+07
36.0	563.8	404.8	0.3155E+06	77	0.2572E+07	0.2798E+07	0.5370E+07
37.0	566.7	407.7	0.3107E+06	75	0.2584E+07	0.2747E+07	0.5331E+07
38.0	569.5	410.5	0.3060E+06	74	0.2597E+07	0.2710E+07	0.5307E+07
39.0	572.4	413.3	0.3015E+06	74	0.2610E+07	0.2691E+07	0.5302E+07
40.0	575.2	416.2	0.2972E+06	73	0.2623E+07	0.2656E+07	0.5279E+07
41.0	578.0	419.0	0.2930E+06	72	0.2636E+07	0.2622E+07	0.5258E+07
42.0	580.9	421.9	0.2891E+06	71	0.2649E+07	0.2589E+07	0.5238E+07
43.0	583.7	424.7	0.2853E+06	70	0.2662E+07	0.2556E+07	0.5219E+07
44.0	586.5	427.5	0.2817E+06	69	0.2675E+07	0.2525E+07	0.5200E+07
45.0	589.4	430.4	0.2782E+06	68	0.2688E+07	0.2494E+07	0.5182E+07
46.0	592.2	433.2	0.2748E+06	67	0.2701E+07	0.2464E+07	0.5165E+07
47.0	595.1	436.0	0.2716E+06	66	0.2714E+07	0.2434E+07	0.5148E+07
48.0	597.9	438.9	0.2684E+06	65	0.2727E+07	0.2404E+07	0.5131E+07
49.0	600.7	441.7	0.2651E+06	65	0.2740E+07	0.2392E+07	0.5132E+07
50.0	603.6	444.6	0.2619E+06	65	0.2753E+07	0.2375E+07	0.5128E+07
51.0	606.4	447.4	0.2587E+06	64	0.2766E+07	0.2346E+07	0.5112E+07
52.0	609.3	450.2	0.2556E+06	64	0.2779E+07	0.2330E+07	0.5109E+07
53.0	612.1	453.1	0.2526E+06	63	0.2792E+07	0.2302E+07	0.5094E+07
54.0	614.9	455.9	0.2497E+06	62	0.2805E+07	0.2274E+07	0.5079E+07
55.0	617.8	458.8	0.2469E+06	62	0.2818E+07	0.2260E+07	0.5078E+07
56.0	620.6	461.6	0.2442E+06	61	0.2831E+07	0.2233E+07	0.5064E+07
57.0	623.5	464.4	0.2416E+06	60	0.2844E+07	0.2206E+07	0.5050E+07
58.0	626.3	467.3	0.2391E+06	60	0.2857E+07	0.2194E+07	0.5051E+07
59.0	629.1	470.1	0.2367E+06	59	0.2870E+07	0.2168E+07	0.5038E+07
60.0	632.0	473.0	0.2343E+06	58	0.2883E+07	0.2143E+07	0.5026E+07
61.0	634.8	475.8	0.2320E+06	58	0.2896E+07	0.2131E+07	0.5027E+07
62.0	637.6	478.6	0.2298E+06	57	0.2909E+07	0.2107E+07	0.5016E+07
63.0	640.5	481.5	0.2276E+06	56	0.2922E+07	0.2083E+07	0.5005E+07
64.0	643.3	484.3	0.2255E+06	56	0.2935E+07	0.2072E+07	0.5007E+07
65.0	646.2	487.1	0.2235E+06	55	0.2948E+07	0.2048E+07	0.4996E+07
66.0	649.0	490.0	0.2215E+06	54	0.2961E+07	0.2025E+07	0.4986E+07
67.0	651.8	492.8	0.2195E+06	54	0.2974E+07	0.2015E+07	0.4989E+07
68.0	654.7	495.7	0.2176E+06	53	0.2987E+07	0.1992E+07	0.4979E+07

	69.0	657.5	498.5	0.2157E+06	53	0.3000E+07	0.1982E+07	0.4982E+07
	70.0	660.4	501.3	0.2139E+06	52	0.3013E+07	0.1960E+07	0.4973E+07
	71.0	663.2	504.2	0.2121E+06	52	0.3026E+07	0.1951E+07	0.4977E+07
	72.0	666.0	507.0	0.2103E+06	51	0.3039E+07	0.1929E+07	0.4968E+07
	73.0	668.9	509.9	0.2086E+06	51	0.3052E+07	0.1920E+07	0.4972E+07
	74.0	671.7	512.7	0.2069E+06	50	0.3065E+07	0.1898E+07	0.4963E+07
	75.0	674.6	515.5	0.2053E+06	50	0.3078E+07	0.1890E+07	0.4968E+07
*	76.0	677.4	518.4	0.2036E+06	49	0.3091E+07	0.1868E+07	0.4959E+07 *
	77.0	680.2	521.2	0.2020E+06	49	0.3104E+07	0.1860E+07	0.4964E+07

- Calculations of Payback Period

$$\text{Payback Period (PBP)} = \frac{\text{Capital Investment}}{\text{Energy Saving}} = \frac{\$\ 256590}{\$\ 259860\ /\text{year}} = \text{One Year}$$

Appendix B

LINGO PROGRAM

Heat integration of the existing heat exchangers network to find minimum utilities requirement using LINGO set formulation;

```
Min = QHmin + QCmin;

R1 - QHmin = -1.14273;
R2 - R1 = -194.147;
R3 - R2 = -0.1;
R4 - R3 = -239.34;
R5 - R4 =-30.71;
R6 - R5 =-2.15;
R7 - R6 =-56.92;
R8 - R7 =-87.16;
R9 - R8 =-4.91;
R10- R9 =-6.87;
R11 - R10 =2.39;
R12 - R11 =13.33;
R13- R12=0.49;
R14 - R13 =1.46;
R15- R14=5.03;
R16- R15 =44.93;
R17-R16=4.728;
R18- R17 =-3.05;
R19 - R18 =-4.44;
R20 - R19 =-0.95;
R21- R20=-4.6;
R22 - R21=2.37;
R23 - R22 =0.66;
R24 - R23 =19.08;
R25- R24=12.8;
R26 - R25 =3.06;
R27 - R26 =1.84;
R28 - R27 =0.3;
R29- R28=3.16;
R30- R29 =-1.46;
R31 - R30 = 0.06;
R32 - R31 = -0.68;
R33- R32=3.74;
R34 - R33 =16.49;
R35 - R34=12.79;
R36 - R35 =9.37;
R37 - R36 =6.65;
R38 - R37 =16.27;
```

```
R39 - R38 =38.87;
R40 - R39 =8.81;
R41 - R40 =1.28;
R42 - R41 =56.6;
R43 - R42 =13.24;
R44 - R43 =6.26;
R45 - R44 =7.21;
R46- R45=7.6;
R47- R46 =3.94;
R48 - R47 =3.44;
R49 - R48=9.1;
R50 - R49=2.82;
R51 - R50=5.36;
R52 - R51 =3.46;
R53 - R52=4.7;
R54 - R53=16.31;
R55 - R54 =98.52;
R56 - R55 =6.47;
R57- R56=0.34;
R58 - R57 =2.27;
QCmin - R58 =2.05;
QHmin >= 0;
QCmin >= 0;
R1 >= 0;
R2 >= 0;
R3 >= 0;
R4 >= 0;
R5 >= 0;
R6 >= 0;
R7 >= 0;
R8 >= 0;
R9 >= 0;
R10 >= 0;
R11 >= 0;
R12 >= 0;
R13 >= 0;
R14>= 0;
R15 >= 0;
R16 >= 0;
R17 >= 0;
R18 >= 0;
R19>= 0;
R20 >= 0;
R21>= 0;
R22 >= 0;
R23 >= 0;
R24 >= 0;
R25 >= 0;
R26 >= 0;
R27 >= 0;
R28 >= 0;
R29 >= 0;
R30 >= 0;
R31 >= 0;
R32 >= 0;
R33 >= 0;
R34 >= 0;
R35 >= 0;
R36 >= 0;
```

```
R37 >= 0;
R38 >= 0;
R39 >= 0;
R40 >= 0;
R41 >= 0;
R42 >= 0;
R43 >= 0;
R44>= 0;
R45 >= 0;
R46 >= 0;
R47 >= 0;
R48 >= 0;
R49>= 0;
R50 >= 0;
R51>= 0;
R52 >= 0;
R53 >= 0;
R54 >= 0;
R55 >= 0;
R56 >= 0;
R57 >= 0;
R58 >= 0;
end
```

```
Global optimal solution found.
  Objective value:                              1087.918
  Infeasibilities:                              0.000000
  Total solver iterations:                             0
```

Variable	Value	Reduced Cost
QHMIN	623.4497	0.000000
QCMIN	464.4680	0.000000
R1	622.3070	0.000000
R2	428.1600	0.000000
R3	428.0600	0.000000
R4	188.7200	0.000000
R5	158.0100	0.000000
R6	155.8600	0.000000
R7	98.94000	0.000000
R8	11.78000	0.000000
R9	6.870000	0.000000
R10	0.000000	2.000000
R11	2.390000	0.000000
R12	15.72000	0.000000
R13	16.21000	0.000000
R14	17.67000	0.000000
R15	22.70000	0.000000
R16	67.63000	0.000000
R17	72.35800	0.000000
R18	69.30800	0.000000
R19	64.86800	0.000000
R20	63.91800	0.000000
R21	59.31800	0.000000
R22	61.68800	0.000000
R23	62.34800	0.000000

R24	81.42800	0.000000
R25	94.22800	0.000000
R26	97.28800	0.000000
R27	99.12800	0.000000
R28	99.42800	0.000000
R29	102.5880	0.000000
R30	101.1280	0.000000
R31	101.1880	0.000000
R32	100.5080	0.000000
R33	104.2480	0.000000
R34	120.7380	0.000000
R35	133.5280	0.000000
R36	142.8980	0.000000
R37	149.5480	0.000000
R38	165.8180	0.000000
R39	204.6880	0.000000
R40	213.4980	0.000000
R41	214.7780	0.000000
R42	271.3780	0.000000
R43	284.6180	0.000000
R44	290.8780	0.000000
R45	298.0880	0.000000
R46	305.6880	0.000000
R47	309.6280	0.000000
R48	313.0680	0.000000
R49	322.1680	0.000000
R50	324.9880	0.000000
R51	330.3480	0.000000
R52	333.8080	0.000000
R53	338.5080	0.000000
R54	354.8180	0.000000
R55	453.3380	0.000000
R56	459.8080	0.000000
R57	460.1480	0.000000
R58	462.4180	0.000000

Row	Slack or Surplus	Dual Price
1	1087.918	-1.000000
2	0.000000	1.000000
3	0.000000	1.000000
4	0.000000	1.000000
5	0.000000	1.000000
6	0.000000	1.000000
7	0.000000	1.000000
8	0.000000	1.000000
9	0.000000	1.000000
10	0.000000	1.000000
11	0.000000	1.000000
12	0.000000	-1.000000
13	0.000000	-1.000000
14	0.000000	-1.000000
15	0.000000	-1.000000
16	0.000000	-1.000000
17	0.000000	-1.000000
18	0.000000	-1.000000
19	0.000000	-1.000000
20	0.000000	-1.000000

21	0.000000	-1.000000
22	0.000000	-1.000000
23	0.000000	-1.000000
24	0.000000	-1.000000
25	0.000000	-1.000000
26	0.000000	-1.000000
27	0.000000	-1.000000
28	0.000000	-1.000000
29	0.000000	-1.000000
30	0.000000	-1.000000
31	0.000000	-1.000000
32	0.000000	-1.000000
33	0.000000	-1.000000
34	0.000000	-1.000000
35	0.000000	-1.000000
36	0.000000	-1.000000
37	0.000000	-1.000000
38	0.000000	-1.000000
39	0.000000	-1.000000
40	0.000000	-1.000000
41	0.000000	-1.000000
42	0.000000	-1.000000
43	0.000000	-1.000000
44	0.000000	-1.000000
45	0.000000	-1.000000
46	0.000000	-1.000000
47	0.000000	-1.000000
48	0.000000	-1.000000
49	0.000000	-1.000000
50	0.000000	-1.000000
51	0.000000	-1.000000
52	0.000000	-1.000000
53	0.000000	-1.000000
54	0.000000	-1.000000
55	0.000000	-1.000000
56	0.000000	-1.000000
57	0.000000	-1.000000
58	0.000000	-1.000000
59	0.000000	-1.000000
60	0.000000	-1.000000
61	623.4497	0.000000
62	464.4680	0.000000
63	622.3070	0.000000
64	428.1600	0.000000
65	428.0600	0.000000
66	188.7200	0.000000
67	158.0100	0.000000
68	155.8600	0.000000
69	98.94000	0.000000
70	11.78000	0.000000
71	6.870000	0.000000
72	0.000000	0.000000
73	2.390000	0.000000
74	15.72000	0.000000
75	16.21000	0.000000
76	17.67000	0.000000
77	22.70000	0.000000

78	67.63000	0.000000
79	72.35800	0.000000
80	69.30800	0.000000
81	64.86800	0.000000
82	63.91800	0.000000
83	59.31800	0.000000
84	61.68800	0.000000
85	62.34800	0.000000
86	81.42800	0.000000
87	94.22800	0.000000
88	97.28800	0.000000
89	99.12800	0.000000
90	99.42800	0.000000
91	102.5880	0.000000
92	101.1280	0.000000
93	101.1880	0.000000
94	100.5080	0.000000
95	104.2480	0.000000
96	120.7380	0.000000
97	133.5280	0.000000
98	142.8980	0.000000
99	149.5480	0.000000
100	165.8180	0.000000
101	204.6880	0.000000
102	213.4980	0.000000
103	214.7780	0.000000
104	271.3780	0.000000
105	284.6180	0.000000
106	290.8780	0.000000
107	298.0880	0.000000
108	305.6880	0.000000
109	309.6280	0.000000
110	313.0680	0.000000
111	322.1680	0.000000
112	324.9880	0.000000
113	330.3480	0.000000
114	333.8080	0.000000
115	338.5080	0.000000
116	354.8180	0.000000
117	453.3380	0.000000
118	459.8080	0.000000
119	460.1480	0.000000
120	462.4180	0.000000

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